Fall 2012

A novel membrane bioreactor configuration for nitrogen removal

Timothy Harrison Canter

Department:

A NOVEL MEMBRANE BIOREACTOR CONFIGURATION FOR NITROGEN REMOVAL

by

TIMOTHY HARRISON CANTER

A THESIS

Presented to the Faculty of the Graduate School of the

MISSOURI UNIVERSITYOF SCIENCE AND TECHNOLOGY

In Partial Fulfillment of the Requirements for the Degree

MASTER OF SCIENCE IN ENVIRONMENTAL ENGINEERING

2011

Approved by

Mark Fitch, Advisor
Joel Burken
Jianmin Wang

ABSTRACT

Population growth in the U.S. and abroad has led to a need for increased treatment of wastewater to limit the effects of additional pollution loadings on bodies of water. Additionally, this need is complicated by limitation on energy inputs. The solution is treatment reactors that are robust, simple to operate, effective in treatment, and energy efficient. This research will compare a traditional membrane bioreactor configuration with a design meant to address these issues.

The proposed configuration used differences in hydraulic head to move liquor from a pre-anoxic zone to the aerobic zone and return a portion of it to the pre-anoxic zone for biological denitrification. The benefits of this design were suspected to be increased nitrogen removal while minimizing energy inputs. This unique configuration consisted of one blower, pumps to induce pressure differentials for membrane operation, and a single mechanical mixer for mixing the anoxic zone and providing hydraulic lift.

Complete data sets for both reactors are presented, and the results from the reactors are compared using data that coincides with operation over the same dates (i.e., summer of 2009 and 2010). Efficacy of the reactors is judged based on their mass and volumetric removal rates. This research resulted in a potentially viable reactor design that increases nitrogen removal over a conventional MBR while reducing the number of pumps needed for operation.

ACKNOWLEDGEMENTS

First and foremost Dr. Mark Fitch should be acknowledged for his help in providing focus, guidance, and support for the completion of this research. He, along with Dr. Joel Burken and Dr. Jianmin Wang, provided excellent resources for planning, execution, and analysis of both the overall project and the data derived from the experiments. Special thanks to Dr. Jianmin Wang for providing the inspiration for the baffled membrane bioreactor design. His unique reactor design – the baffled bioreactor – served as the basis for using baffles and weirs to redirect and control flow between zones. Dr. Cesar Mendoza's assistance explaining hydrological theory and helping to develop the estimates for baffle depths, weir heights, and notch angles was critical for reactor design.

Design and fabrication would have been impossible without the help of Gary Abbott (general mechanical assistance), Steve Gabel (fabrication and materials support) and Brian Swift (programming for permeate/backwash control box). Guoqiang Liu deserves a really big thinks for helping operate the reactor and providing direction. Ashwin Krishnamurthy provided assistance when needed during the summer of 2010. Thanks to the City of Rolla for allowing us to use the space at the Southeast WWTP to test the reactors, to Allen McNece at the Southeast WWTP for operational support, and to Leonard Wood institute for funding the initial research.

TABLE OF CONTENTS

LIST OF ILLUSTRATIONS

LIST OF TABLES

1. INTRODUCTION

Population growth in the U.S. and abroad has led to a need for increased treatment of wastewater to limit the effects of additional pollution loadings on bodies of water. Additionally, this need is complicated by limitation on energy inputs. The solution is treatment reactors that are robust, simple to operate, effective in treatment, and energy efficient. This research attempted to provide a solution to this problem by comparing a traditional membrane bioreactor (MBR) to one that was designed with the intention of removing the need for pumps to move liquor between zones in order to improve biological nitrogen removal.

In the winter of 2009 construction began on a bench-scale MBR that was to serve as the control variable for a larger experiment. The bMBR was designed in the fall of 2009, fabricated during the first quarter of 2010, and operated from May through September of the same year. Its purpose was to increase removal rates of nitrogen over the more conventional bench-scale MBR operated in 2009.

The bMBR's key design differentiation is the use of a single anoxic mixer for two purposes. A single-shaft vertical mixer with two propellers was used to (a) provide mixing to the anoxic zone and (b) to provide hydraulic lift to a portion of the liquor in the anoxic zone. The hydraulic lift created by the mechanical mixer may be interpreted as an artificial increase in hydraulic head, which allows liquid to pass over an adjustable weir and into the aerobic zone. Since this increase in anoxic head is induced, the hydraulic head in the aerobic zone is

actually greater than that of the anoxic zone. This difference induces return flow from the aerobic zone to the anoxic zone. Increasing or decreasing the depth of a baffle between the aerobic and anoxic zones controls the rate of this flow.

Overall the experiment seems to have been a success based on volumetric removal and removal rates of nitrogen species of the MBR in 2009 and the bMBR in 2010. However, variances in ambient conditions and municipal influent characteristics could have contributed to the difference in data. It is unknown whether these differences between 2009 and 2010 would have positively or negatively affected the operation of the two reactors (i.e., Would the MBR have been more effective in removing nitrogen had it been operated in 2010?). Certainly more research is needed to assess efficiency and effectiveness of the bMBR as well as the implications of scaling the system.

2. LITERATURE REVIEW

2.1. INTRODUCTION

The advent of the 21st century has seen more stringent regulation to maintain water quality in the United States and has led to increased complexity of wastewater treatment. Biological oxygen demand (BOD)/chemical oxygen demand (COD), ammonia (NH_3), nitrate (NO_3^-)/nitrite (NO_2^-), phosphate (PO_4^{3-}) and pathogens are all considered undesirable effluent constituents. Simple, conventional configurations for biological treatment (trickling filters, activated sludge, etc.) are giving way to configurations that often include multiple liquor recycle pumps, tertiary and advanced treatment. These new, advanced configurations often require a greater amount of energy and man-hours to operate.

The use of a micro- or nano-porous membranes for liquid/solid separation are becoming increasingly popular due to their ability to produce high-quality effluent on a relatively small operational footprint, primarily achieved by eliminating secondary clarification. Biological nutrient removal in most conventional membrane bioreactors (MBRs) still calls for multiple pumps, anoxic mixers and careful flow monitoring and control to optimize reactor efficiency and effectiveness. Notable exceptions include simultaneous nitrification and denitrification (which may be difficult to control on a consistent basis), intermittent

aeration to cycle aerobic/anoxic conditions in the reactor (which creates a batch-type reactor and decreases overall flow rate), and the unique configuration of stacking the aerobic MBR on top of the anoxic zone developed by Kimura et al. (Kimura, Nishisako et al. 2008; Sarioglu, Insel et al. 2009; Yang 2009).

Much of the resistance to widespread adoption of membrane technology is due to its inherent costs, both capital and operational. Capital costs have decreased thanks to improvements in materials technology and membrane module capacities (i.e., higher surface area to volume ratio). Operational costs, primarily related to fouling control, have decreased as operational understanding of membranes have increased (Buer and Cumin 2008). Although fouling of membranes continues to represent a significant hurdle to their widespread application, this research was not focused on membrane fouling and its mediation, due in large part to the extensive amount of literature that already exists on the subject.

The purpose of this research was to explore a unique configuration of a MBR for advanced nitrogen removal that reduces the necessity for pumps to control flow from the anoxic zone to the aerobic zone and recycle flow from the aerobic back to the anoxic. Eliminating these pumps has the potential to save capital and operational costs for future designs, as well as decrease energy requirements for system operation.

2.2. EFFECTS AND REMOVAL OF NITROGEN IN WASTEWATER

One of the primary nutrients of concern in wastewater is nitrogen which, even though required for cellular synthesis and life, can be problematic when it is no longer the limiting factor for biomass growth that leads to eutrophication (Grady Jr., Daigger et al. 1999). Urea from humans is readily converted to ammonia once it has left the body and comes in contact with water. Phosphoric compounds, although often associated with detergents and personal hygiene products, can be found in biological wastes as well. Once these nutrients enter the environment they have the potential to promote eutrophication of receiving bodies of water by encouraging algal growth.

The ratio of oxygen to nitrogen by weight needed to oxidize ammonia to nitrate is 4.33 g O_2/g N (Werzernak 1967). Although organic carbon and phosphoric compounds were traditionally suspected of being responsible for eutrophication, recent studies suggest that nitrogenous compounds may play a large role as well – particularly in estuaries and marine environments (Howarth and Marino 2006).

Nitrogen in the environment is seen in several forms, and some portion of the total nitrogen in the environment is in flux between forms. The entire system of species and fluxes is often referred to as the Nitrogen Cycle. The portion of the Nitrogen Cycle that wastewater treatment concerns itself with begins with ammonia and ends with nitrogen gas and/or cellular matter. The first step of nitrification is the oxidation of ammonia to nitrite, typically attributed to

Nitrosomonas. In the next step, *Nitrobacter* oxidizes nitrite to form nitrate (Grady Jr., Daigger et al. 1999). The stoichiometry of biological nitrification is represented from ammonia to nitrite (Equation 1), from nitrite to nitrate (Equation 2), and, in total, from ammonia to nitrate (Equation 3). Combining half reactions gives the overall nitrification reaction, shown in Equation 4 (Tchobanoglous, Burton et al. 2003). This equation suggests that 98% of nitrified nitrogen is oxidized to nitrate, while 2% is converted to biomass.

$$2NH_4^+ + 3O_2 \longrightarrow 2NO_2^- + 4H^+ + 2H_2O \quad (1)$$

$$2NO_2^- + O_2 \longrightarrow 2NO_3^- \quad (2)$$

$$NH_4^+ + 2O_2 \longrightarrow NO_3^- + 2H^+ + H_2O \quad (3)$$

$$NH_4^+ + 1.863O_2 + 0.098CO_2 \longrightarrow 0.0196C_5H_7NO_2 + 0.98NO_3^- + 0.0941H_2O + 1.98H^+ \quad (4)$$

While nitrate isn't seen as strongly contributing to eutrophication, its presence in significant concentrations (greater than 10 mg/L) is potentially hazardous to human health. Most notably it is known to cause "blue baby syndrome" in which nitrate displaces oxygen on red blood cells, giving the affected baby a bluish appearance (Wisconsin Department of Natural Resources 2003).

Both anoxic and anaerobic conditions refer to lack of oxygen. In environmental engineering the term "anoxic" specifically refers to the presence of

nitrate or nitrite in the surrounding environment. Therefore, any future reference in this paper using the term anaerobic refers to situations in which the environment lacks both oxygen and nitrate or nitrite.

The next step of the cycle – denitrification – occurs when nitrate, rather than oxygen, is utilized as an electron acceptor by bacteria surviving under anoxic conditions. As with nitrification, denitrification involves multiple steps including the reduction of nitrate to nitric oxide, nitrous oxide, and nitrogen gas (Tchobanoglous, Burton et al. 2003). The formation of molecular nitrogen, which is typically off-gassed into the atmosphere, is accompanied by a small amount of ammonia, carbon dioxide, water, and hydroxide ions. The stoichiometry for denitrification using biodegradable organic matter in wastewater as the electron donor is noted in Equation 4:

$$C_{10}H_{19}O_3N + 10NO_3^- \longrightarrow 5N_2 + 10CO_2 + 3H_2O + NH_3 + 10OH^- \quad (5)$$

Several species of bacteria, both heterotrophic and autotrophic, are capable of denitrification. However, most are considered to be facultative aerobic organisms with the ability to use nitrate as well as oxygen (Gayle 1989). *Pseudomonas* is one specific heterotrophic genus considered by many to be the most common and widely distributed of all the denitrifiers, and such

Pseudomonades have been shown to use a wide array of organic compounds as an electron donor (Payne 1981).

2.3. BACKGROUND ON MEMBRANES FOR WASTEWATER TREATMENT

Advancements in materials and system configurations have increased the popularity of membrane technologies in recent years. Although membranes have been used to some degree in water treatment systems since the 1960s to physically separate water from particulates (Stephenson 2000), the technology has only become more widespread in recent years. Reduction of membrane material cost and operational cost, as well as more stringent regulation that has increased the operational cost of other methods, has made membrane technology much more cost-competitive (Daigger 2005).

Membrane configurations have taken two primary forms – submerged or immersed MBRs and sidestream MBRs. The immersed membrane configuration is slightly more recent and allows for greater membrane surface area within a comparable volume. Furthermore, sidestream membrane configurations must be operated under higher transmembrane pressures, resulting in greater fouling potential and, as a result, greater demand for energy and maintenance (Judd 2006). In fact, it has been reported that immersed membranes can offer biosolids separation for a mere 20% of the energy requirements of sidestream membranes (Buer and Cumin 2008). On the other hand, proponents of

sidestream membrane configurations claim their systems are easier to retrofit into older treatment plants with membrane technology (Stephenson 2000).

Two membrane configurations that are relatively new and, therefore, relatively untested are the extractive and diffusive membrane configurations. While we typically think of membranes as retaining biomass and solids while allowing relatively clean liquid permeate to pass, the extractive membrane is designed to extract specific compounds or constituents from the water as permeate. These extracts are then treated separately from the rest of the liquor. The diffusive membrane is used to introduce molecular, or “bubbleless”, gas into a reactor, usually to support a biofilm (Judd 2006).

When either sidestream or immersed membranes are combined with an aerobic, suspended-growth biological reactor, they replace the clarifier and thus also the hydraulic restraints placed upon system operation due to the overflow limitations of the clarifier. Thus, MBRs allow for almost complete independent control over sludge retention time (SRT) and hydraulic retention time (HRT) (Melin 2006).

Since the membrane replaces the clarifier in a conventional system, the footprint of the facility is decreased and the biomass may be retained in the reactor without additional sludge return. The latter feature allows for mixed liquor suspended solids (MLSS) concentrations in membrane bioreactors to become much higher than in conventional systems – up to 25 g/L – although operational efficiencies limit MLSS concentrations to 8-10 g/L (Metcalf & Eddy, Tchobanoglous et al. 2003; Tchobanoglous, Burton et al. 2003). Higher MLSS

concentrations allow MBRs to provide highly efficient (in terms of volume) removal of constituents, particularly for high-strength wastewater, as well as an acute ability to acclimate shock loads. Furthermore, microfiltration, with nominal pore sizes of 0.08-0.2 μm, serves to provide a considerable amount of disinfection by acting as a barrier for bacteria.

Despite all their advantages, MBRs are considered to have some disadvantages associated with them as well. These include relatively high capital and maintenance costs, requirements for operators with high levels of technical expertise, high-energy consumption rates, and constant concern over the wellbeing of the membrane due to pressure, pH, large particulates and destructive chemicals in the influent. Also, the treatability of the wasted sludge is questionable (Melin 2006).

Membrane flux may be the single most important parameter of membrane operation. Membrane flux is expressed as volume of permeate per membrane area and time. For example, metric units for this may be m^3/m^2*sec, which may be simplified to m/sec and referred to as permeate velocity (Stephenson 2000).

The concept of critical flux was first proposed several years ago, and it is defined as “the flux below which an increase in transmembrane pressure (TMP) does not occur with time” (Field 1995). An example of this may be seen in Figure 2.1 where the arrow designates the point of critical flux.

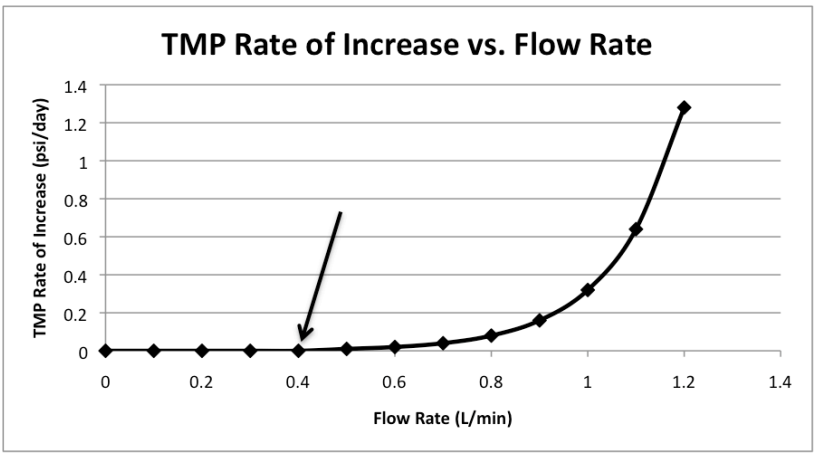


Figure 2.1. Illustrative Example of Critical Flux

In Figure 2.1 the flux is represented by flow rate as the two are interchangeable when referring to a single membrane unit. Below the critical flux the membrane is operating at an inefficient flow rate or flux. Above the critical flux membrane fouling begins to occur, in this case defined as rate of TMP increase.

While this concept is novel and theoretically applied, it has been widely reported that TMP still increases when operational flux is below critical flux. The latter is more intuitive. If a membrane were placed in water with no flux, biofilm would accumulate on the membrane surface and, ultimately, lead to fouling. In operational settings the TMP increases as a function of the flux applied to the membrane as seen in Figure 2.2.

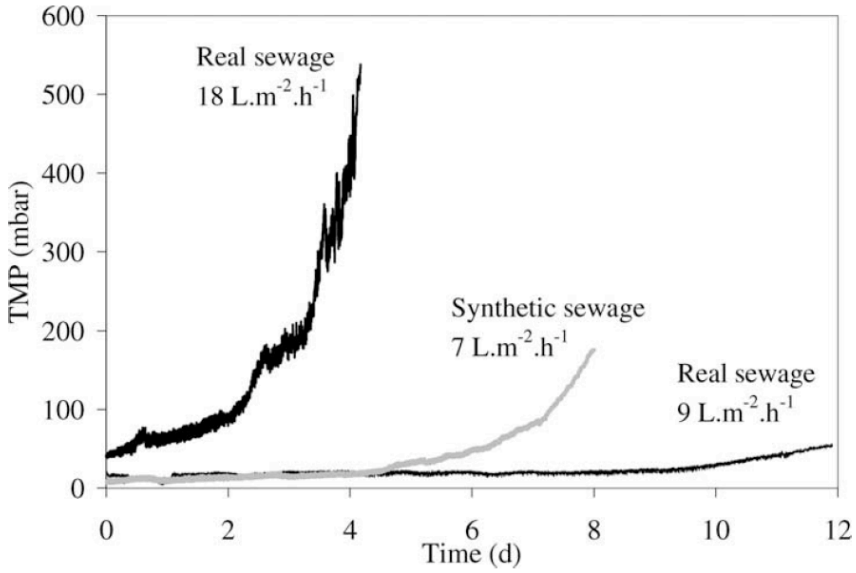


Figure 2.2. Changes in TMP with Differing Flux in Submerged MBR (Le Clech 2003)

Wang et. al. assessed flux sustainability in a study where flux lower than critical flux – referred to as sub-critical flux – had been applied (Wang 2008). Figure 2.3 shows the results of their experiments with an operational flux of 25 L/(m^2*h) and a critical flux of 32-38 L/(m^2*h). The arrows in Figure 2.3 denote the times when chemical cleaning of the membrane was necessary. The author did not provide explanation or hypothesis for the difference in operational timeframes between cleaning. It could be that the differences between recovery cleanings were the result of the characteristics of the influent. Higher concentrations of specific constituents (e.g., fats, oils, grease, heavy metals, etc.) may clog the membrane more quickly.

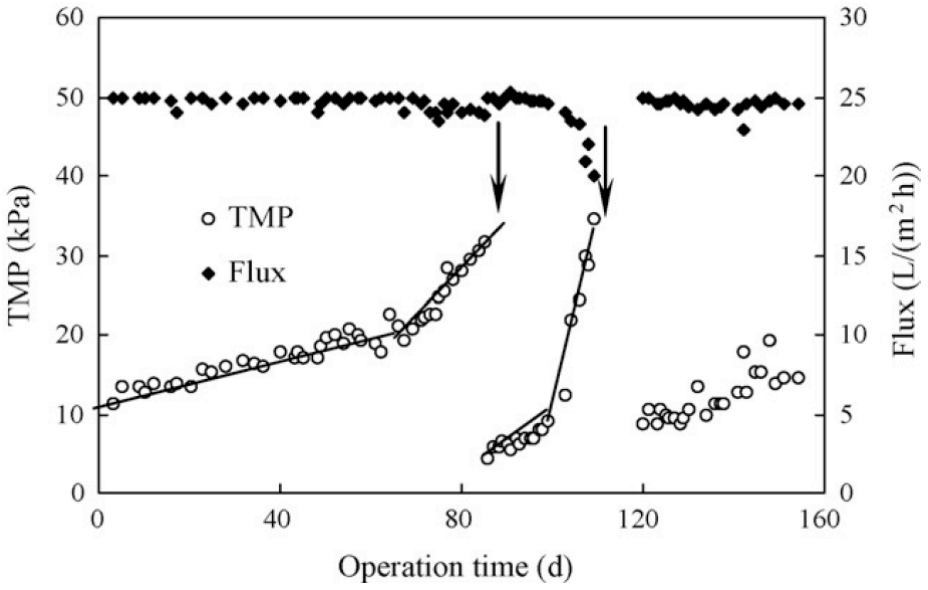


Figure 2.3. TMP Changes Under Sub-Critical Flux Operation (Wang 2008)

Results seem to indicate that membrane-fouling characteristics are seen as being somewhat rapid before reaching a phase of sustainable operation for a relatively long period of time. This is followed by a sharp increase in TMP once a significant amount of irreversible fouling has occurred. Furthermore, membrane operation below sub-critical flux seems to control caking of biomass on the membrane surface, but TMP still increases due to a gel-like layer that's believed to be composed of organic macro-molecules such as soluble microbial products (SMP), extracellular polymeric substances (EPS) and other substances released during cell lysis (Wang 2008).

Fouling of the membrane in MBRs is the root cause of increase in TMP. The rate of fouling, flux aside, is dependent on three factors: biomass characterization (MLSS concentration, extracellular polymeric substances,

flocculation, etc.), operational conditions (aeration scouring, permeate flow rate, backwash flow rate and frequency, etc.) and the membrane's physical and chemical characteristics (pore size, hydrophilicity/hydrophobicity, surface charge, etc.) (Yigit 2008).

Although much of the literature speaks of fouling in general, it is commonly subcategorized. Traditional differentiation has focused on reversible fouling (that which can be removed) and irreversible fouling (that which can't be removed). However, since these terms were introduced, their precise meanings have become convoluted as some fouling may be removed through physical means (air scouring and backwashing) or chemical means (acid, base or bleach treatments) while some fouling is permanent. Therefore, it's preferable to characterize these three types of fouling as removable (physical), irremovable (must chemically treat to remove), and irreversible (membrane is permanently affected).

Removable fouling is typically attributed to biocaking and soluble microbial product (SMP) accumulation on the membrane's surface. Irremovable fouling is predominated by the accumulation of SMP and various colloids and solutes on the membrane surface as well as within the membrane's pores, but may also include inorganic fouling from the precipitation of metals and salts from the solute. Irreversible fouling is typically attributed to inorganic fouling as some inorganic substances – particularly metals – may physically and/or chemically bond to inorganic membranes. (Meng, Chae et al. 2009). An illustrative difference between these three types of fouling can be seen in Figure 2.4.

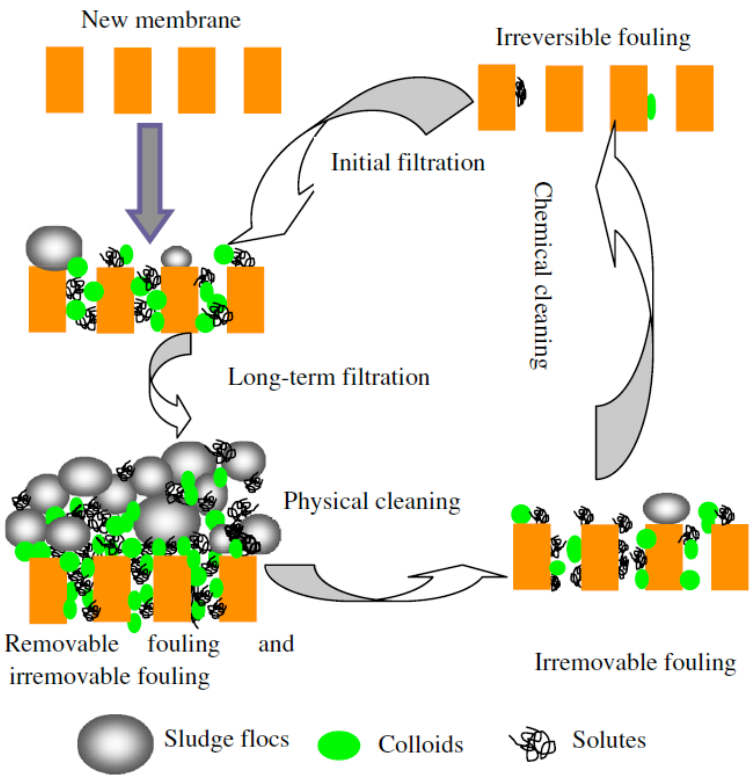


Figure 2.4. Removable, Irremovable and Irreversible Fouling (Meng, Chae et al. 2009)

With flux, TMP, fouling and their interrelationship being critical to the cost of operation for MBRs, much research has been devoted to reducing the rate of fouling and TMP increase while maintaining cross-membrane flux. It is well understood that the vast majority of fouling that occurs can be removed by chemical treatment of acid, base, and bleaching agents. However, such chemical treatments can decrease the membrane's lifespan, are costly, and have the potential to cause environmental problems (Meng, Chae et al. 2009). Therefore, it makes sense to control fouling through preventative means.

Generally speaking, there are three types of preventative fouling controls for membrane operation: physical scouring of the membrane using coarse-bubble aeration, backwash cycles to reverse advective transport and flush particles within the membrane pores (then carried away by bubble scouring), and modification of operational parameters to affect sludge/liquor characteristics (e.g., solids retention time (SRT)).

As the permeate cycle of membrane operation progresses, solutes and colloids are entrapped in the pores of the membrane. Providing an intermittent backwashing cycle has been found to successfully remove a significant amount of pore blocking and partially remove or loosen caking that has occurred on the membrane's surface – allowing scouring by aeration to more easily remove the foulants (Le-Cletch 2006).

When one thinks about different backwash scenarios it is somewhat intuitive that as backwash duration and frequency are increased the rate of fouling decreases. While true, there is a point where too long a backwash duration or too frequent a backwash cycle causes inefficient operation. Some literature suggests that optimal operation, in terms of fouling control and net daily permeate volume, is roughly 9:45-10 minutes of permeate cycle to 5-15 sec of backwash – so long as operational flux was below critical flux (Yigit 2008). Others have noted that less frequent, longer backwashing cycles were more efficient at removing foulants than frequent backwashing cycles of short duration. Less frequent, longer backwashing cycles may encourage membrane layer caking, which in turn prevents small particles, such as colloids, from entering the

pores of the membrane (Jiang 2003). This may be seen in Figure 2.5 where cake filtration does not seem to set in until after 100 seconds of permeate cycle.

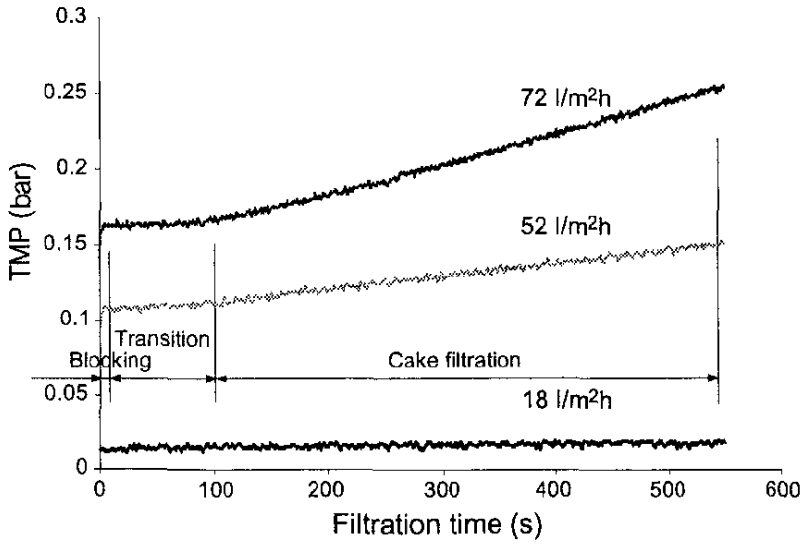


Figure 2.5. Membrane Filtration Types After Backwash Cycle (Jiang 2003)

Unlike traditional aerobic wastewater treatment reactors that typically use fine-bubble aeration to mix and oxygenate the liquor, membrane bioreactors generally use coarse-bubble aeration to provide oxygen and scour caking from the membrane surface. The scouring of the caking improves membrane flux (Psoch 2006a). Although small bubbles are more efficient at transferring oxygen to the surrounding liquor, large bubbles are preferable for controlling membrane fouling (Phattaranawik 2007). More than one study has suggested that increased aeration intensity leads to increased permeability and decreased

fouling rate (Trussell 2007; Nywening and Zhou 2009). Air sparging of membranes usually occurs from the bottom of the membrane, allowing the bubbles to scour from the bottom up. At least one study has shown that placing diffusers on both the top and bottom of the membrane unit seems to slow the increase of transmembrane pressure while decreasing the loss of permeability (Park, Lee et al. 2010). The amount of aeration provided to a membrane is limited by the cost of power use and restrictions placed on operators by membrane manufacturers to maintain membrane integrity (Judd 2006).

The operational parameters of SRT and HRT can be modified to provide conditions under which the operational life of the membrane is increased. While these parameters do not directly affect membrane permeability, flux or transmembrane pressure, they do directly affect sludge characteristics that, in turn, affect the aforementioned membrane factors. Several authors have concluded that decreases in HRT seem to increase membrane fouling (Cho 2005; Chae 2006) (Meng 2007). Both Chae and Meng agreed that a decrease in HRT lead to increased concentrations of EPS and larger particles that had a negative affect on membrane permeability.

Like HRT, a decrease in SRT seems to increase fouling. Studies have indicated that a SRT decrease from 100 to 20 days, 30 to 10 days, and from 5 to 3 days all suggested the same trend of increased transmembrane pressure and fouling (Ng 2006; Zhang 2006; Ahmed 2007). Unlike HRT, modifications to SRT seemed to primarily affect SMP concentrations as opposed to EPS concentrations. As SRT increases, SMP concentrations increase leading to

increased fouling potential (Liang 2007). It should be noted that some studies have found increased fouling rates associated with prolonged SRT. The authors credit this phenomenon to increased concentrations of foulants (they did not define what these foulants were) and high fluid viscosity while noting critical flux decreased (Han 2005).

2.4. PRINCIPLES OF NITROGEN REMOVAL IN CONVENTIONAL SYSTEMS

Development of different aquatic environments to select for the growth of microorganisms allows for treatment of water in terms of hours rather than the days or weeks it would take if the water were directly discharged into the environment and natural methods were allowed to take their course. The use of microbes to degrade organic carbon and nutrients is ideal for wastewater treatment as they are relatively inexpensive. Aerobic and anoxic zones offer the environmental conditions considered necessary for nitrogen removal.

There are two primary configurations through which nitrogen is removed in wastewater treatment. The first is generally termed a “step system.” Such systems may come in different forms: completely mixed, plug-flow, or batch reactors. They are defined by separate zones that have specific purposes – although these zones may not necessarily be divided by physical partitions. For the purposes of this discussion and research we will only consider suspended-growth configurations designed with a preanoxic zone. Postanoxic zone

configurations place the anoxic zone downstream from the aerobic zone and require an exogenous carbon source or are dependant on endogenous decay from the aerobic zone for organic carbon. If the latter is the case, the denitrification rates in postanoxic designs is typically slower than other designs by a factor of 3 to 8 (Tchobanoglous, Burton et al. 2003).

In preanoxic configurations, influent is first introduced to the anoxic zone where the organic carbon is utilized as the electron donor during denitrification. The liquor leaves the anoxic zone and enters the aerobic zone where nitrification occurs and organic carbon is further assimilated. Some effluent is allowed to leave the aerobic zone and continue to the next step of the treatment process. Additionally, some effluent is recycled back to the beginning of the anoxic zone where it's mixed with influent to provide nitrate to the anoxic zone to keep it from reaching anaerobic conditions. Most completely mixed activated sludge designs for biological nutrient removal operate in this manner and are separated by partitions. Plug flow reactors operate on the same principle, but without physical partitions. Batch reactors may employ intermittent aeration to achieve the same functions, but the zone is essentially separated chronologically rather than spatially. That is, even though anoxic and anaerobic zones are created, they occur in the same volume but during different periods (Tchobanoglous, Burton et al. 2003).

Another type of configuration encourages simultaneous nitrification and denitrification within the same volume and during the same period of time. Two specific designs have seemed to achieve this feat – oxidation ditches and

membrane bioreactors. Oxidation ditches that are operated with relatively low D.O. concentrations and long SRTs have seen high levels of both nitrification and denitrification. These results were due to the unique properties of flocculating bacteria, insomuch as the cells on the outside of the flocs are able to use the oxygen before it can reach the internal cells. Thus, the outside cells are responsible for nitrification and the internal cells for denitrification (Tchobanoglous, Burton et al. 2003). Simultaneous nitrification/denitrification has also been noted in membrane bioreactors with the same floc-diffusion principle thought to be responsible (Sarioglu, Insel et al. 2009).

Table 2.1 displays a comparison of effluent values for a number of different processes.

Table 2.1. Process Performance Comparisons (Patel 2005)

Process	**COD:TKN Ratio**	**HRT (h)**	**SRT (day)**	**Eff. COD (mg/L)**	**Eff. NH_4-N (mg/L)**	**Eff. NO_3-N (mg/L)**	**Eff. TKN (mg/L)**
A2O	100:13.6	10	10	–	7.5	4.5	10.5
UCT	100:10.7	17.5	20	29	–	9.6	4.0
MUCT	100:9.4	17.5	20	35	–	9.3	4.8
MBR (Single Stage)	100:12	7.5	20	17	<0.3	43	<0.5
Modified MBR	100:9.4	12	20	3	0.4	5.8	–

"Single Stage MBRs" refers to aerobic MBRs with no connecting anoxic or anaerobic chamber. The "Modified MBR" they refer to consists of an anaerobic zone, clarifier, anoxic zone and aerobic zone with a submerged membrane as seen in Figure 2.6 (Patel 2005).

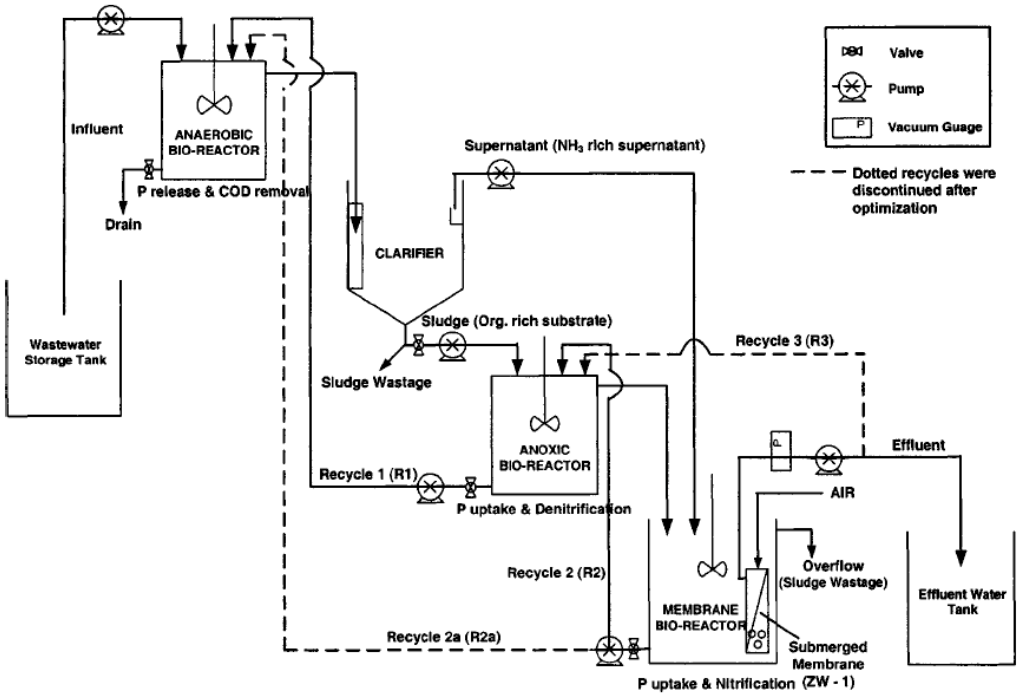


Figure 2.6. Modified Membrane Bioreactor (Patel 2005)

2.5. NOVEL MEMBRANE BIOREACTOR CONFIGURATIONS FOR ADVANCED NITROGEN REMOVAL

The "Modified MBR" noted in the last section and displayed in Figure 2.6 represent one of several different configurations that have been employed to

effectively remove nutrients from wastewater. Creators of these different configurations appear to have had different goals in mind – be they most effective constituent removal or simplicity and efficiency of operation – and their creations may be categorized as such. For the purposes of our discussion we'll separate the configurations into the following groups:

- Systems with separate anoxic or anaerobic zones in separate tanks that require pumps for moving liquor and return sludge between them.
- Systems that employ two different reactors (i.e., packed-bed or biofilm) in addition to the MBR for nutrient removal.
- Systems that separate aerobic from anoxic and/or anaerobic using baffles or hydraulics.

The first two listed configurations include reactors designed for maximum constituent removal. More often than not, these designs are complex in configuration and operation; arguably requiring increased maintenance, construction footprint and capital cost. However, their ability to remove nitrogen, phosphorous and organic carbon from waste streams is very good – making them ideal for installations where effluent water quality is of the highest concern.

The third type of configuration is more representative of compact systems that may or may not provide extremely high effluent quality, but which are simple in their design and operation – allowing for minimal capital and operational costs. These types of systems seem to be ideal for smaller, perhaps less affluent communities who seek to meet environmental regulations with the least economic impact on citizens.

Configurations not considered are those that require methanol, alum or organic flocculent addition to achieve additional nutrient removal. Additions of such substances have been proven effective in conventional systems as well as in differing MBR configurations (Fleischer 2005; Song 2008; Ngo and Guo 2009). Therefore, dosing reactors with different substances are viewed as enhancements to configurations rather than unique configurations in and of themselves.

2.5.1. Systems With Anoxic/Aerobic Zones Completely Separated. These types of systems are more reminiscent of conventional treatment systems whereby forward flow is induced via hydraulic head gradient and liquor/sludge is returned via mechanical pump. The primary advantage of such systems is that they allow for maximum control over environmental control in each section of the reactor. By separating the anoxic zone from the aerobic zone one can better ensure minimal D.O. in the anoxic zone. The same can be said for nitrate/nitrite in the anaerobic zone with an anaerobic/anoxic comparison. While well-defined separation of the zones allows for more precise control, such systems also require a considerable footprint for construction. Therefore, the primary disadvantage is increased capital cost.

One such example of this type of system can be seen in Figure 2.6. Patel et. al. designed a system that is complex in terms of operation and layout (Patel 2005). As such, capital costs and operational and maintenance (O&M) costs are assumably quite high. However, the high-quality effluent these researchers were

able to produce may justify these expenses. Effluent of this system is noted as "Modified MBR" in Table 2.1.

A second, albeit simplified, example of a separated configuration is provided by Brannock et. al. (Brannock, Leslie et al. 2009). Figure 2.7 shows this configuration with a flat sheet membrane being used in the design on the left and a hollow-fiber membrane being used in the design on the right. While both designs use mechanical pumps for liquor return, the cyclical design on the right reduces capital cost by maximizing the number of shared walls.

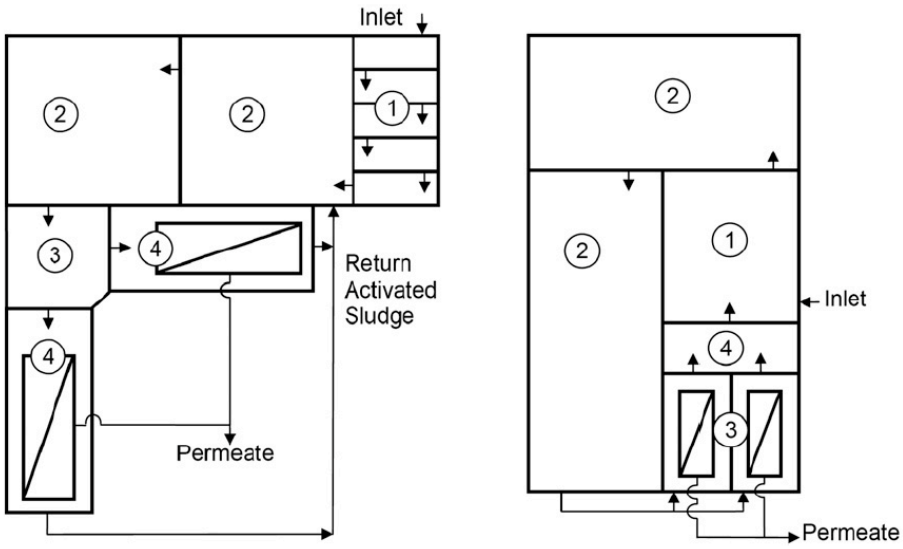


Figure 2.7. Separated Membrane Bioreactor Configuration (Brannock, Leslie et al. 2009)

In the reactor on the left, influent enters a bioselector (1) before flowing to swing aerobic and anoxic zones (2). From there the liquor flows to an aerobic

zone (3) and, finally, to membrane filtration (4). In the reactor on the left the influent is introduced to an anoxic zone (1) followed by aerobic zones (2) and into membrane filtration (3). Some liquor is recycled through a deaeration zone (4) before reentering the anoxic zone (1). Table 2.2 shows effluent data for these designs, with Site 1 referring to the design on the left and Site 2 referring to the design on the right.

Table 2.2. Effluent Values of MBR Configurations in Figure 2.7 (Brannock, Leslie et al. 2009)

Site	Eff. COD (mg/L)	Eff. NH_3-N (mg/L)	Eff. NO_x-N (mg/L)	TSS (mg/L)
Site 1	48.8	0.7	1.5	1.5
Site 2	29.0	0.1	16.1	1.0

2.5.2. Systems that Employ Hybrid Reactors. Systems that combine traditional or unique suspended growth reactor configurations with a MBR and media for encouraging attached growth are becoming increasing common. Generally, these types of reactors allow overall footprint reduction by concentrating active biomass in specific sections of the reactor to help encourage nutrient removal.

The first example of a hybrid reactor is the combination of a membrane bioreactor with a membrane biofilm reactor (Figure 2.8). This reactor was configured for the expressed purpose of maximizing nitrogen removal. In this configuration the MBR is used for organic oxidation and nitrification, and the

membrane biofilm reactor is used for denitrification with molecular hydrogen gas being utilized as the electron donor rather than organic carbon (Hasar 2009).

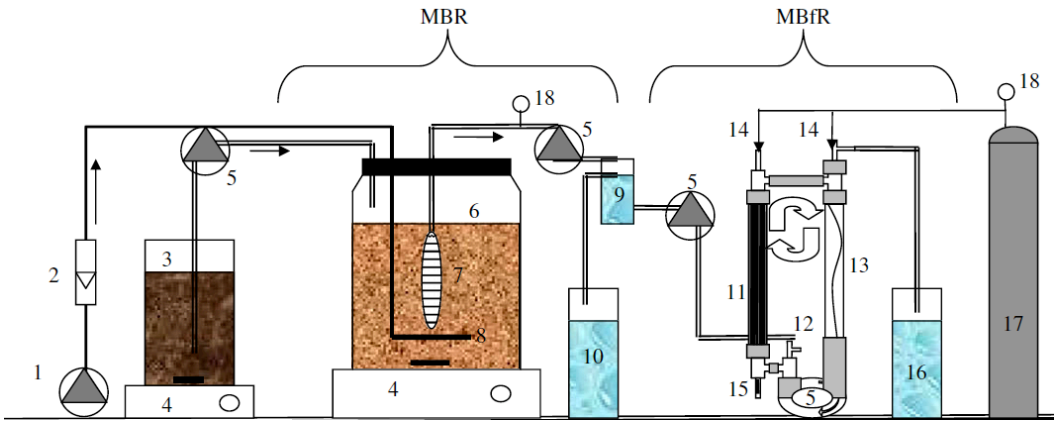


Figure 2.8. MBR Combined with Membrane Biofilm Reactor (Hasar 2009)

Figure 2.8 shows the innate complexity of the system with influent tank (3) leading to the aerobic MBR (7) and then on to the bubbleless gas transfer membrane (11). While this system operates on a reduced footprint, the O&M costs appear to be significant with two membrane systems and the use of hydrogen gas. The surface loading, flux and percent removal of total nitrogen and nitrate with respect to time is illustrated in Figure 2.9, with (A) representing total nitrogen and (B) representing nitrate-nitrogen.

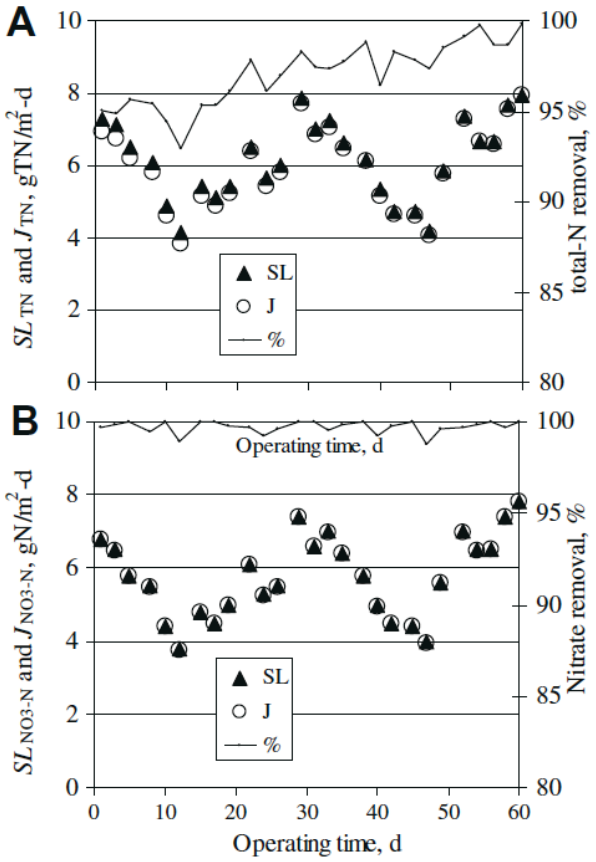


Figure 2.9. Surface Loading, Flux and Percent Removal of Total Nitrogen and Nitrate-Nitrogen in the MBR/Membrane Biofilm Reactor (Hasar 2009)

The second example of a hybrid system is the sequencing batch moving bed membrane bioreactor. The operational theory behind this reactor is that media can be introduced into the aerobic zone of a conventional membrane bioreactor to provide surface area for attached growth organisms. Once firmly established, the underlying layer of biofilm is induced into anoxic conditions by limited oxygen diffusion through the outer layer. Thus, denitrification is accomplished in the aerobic zone (Yang 2009). Figure 2.10 shows the configuration of this reactor.

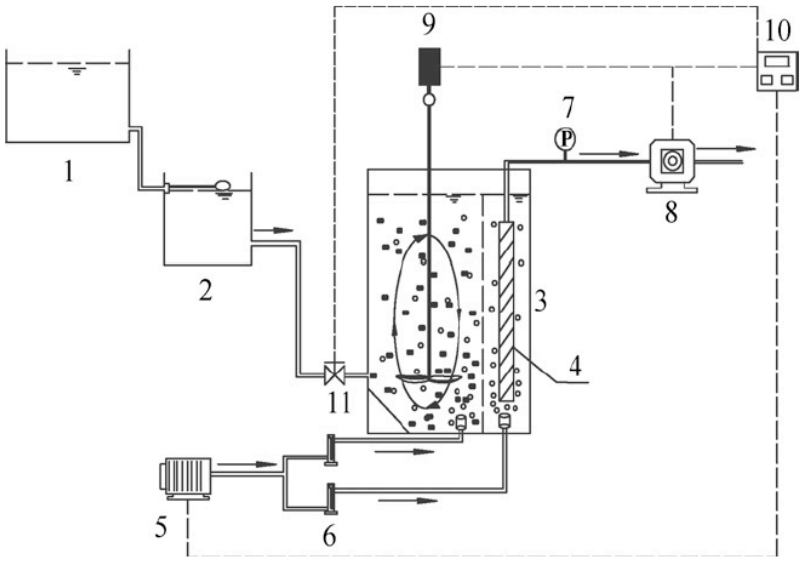


Figure 2.10. Sequencing Batch Moving Bed Membrane Bioreactor (Yang 2009)

This reactor has a reduced footprint, and should be relatively low in O&M costs due to both small footprint and simple operation. The batch nature of this reactor's operation would necessitate a redundant reactor and/or an equalization tank of significant size. Furthermore, inconsistent thickness of biofilm may contribute to variances of effluent quality over time as the active denitrifying biomass changes due to sloughing and/or influent loading. It is assumed that there is physical separation between the media in the moving bed and the membrane in the form of a coarse screen to keep media from damaging the membrane. Otherwise, extended operation of such a system may reduce the operational life of the membrane. Table 2.3 shows removal efficiencies of the sequencing batch moving bed bioreactor demonstrated in Figure 2.10.

Table 2.3. Experimental Results of Sequencing Batch Moving Bed Bioreactor (adapted from (Yang 2009))

Constituent	Average Removal Efficiency (%)
COD	93.5
NH_4-N	95.4
TN	76.9

A membrane bioreactor combined with a packed-bed biofilm reactor constitutes our third hybrid reactor example. As seen in Figure 2.11, it employs media for biofilm growth post-MBR like the membrane biofilm reactor but uses an external carbon source (trisodium citrate) rather than hydrogen gas as the electron donor (Zhang, Zhou et al. 2009).

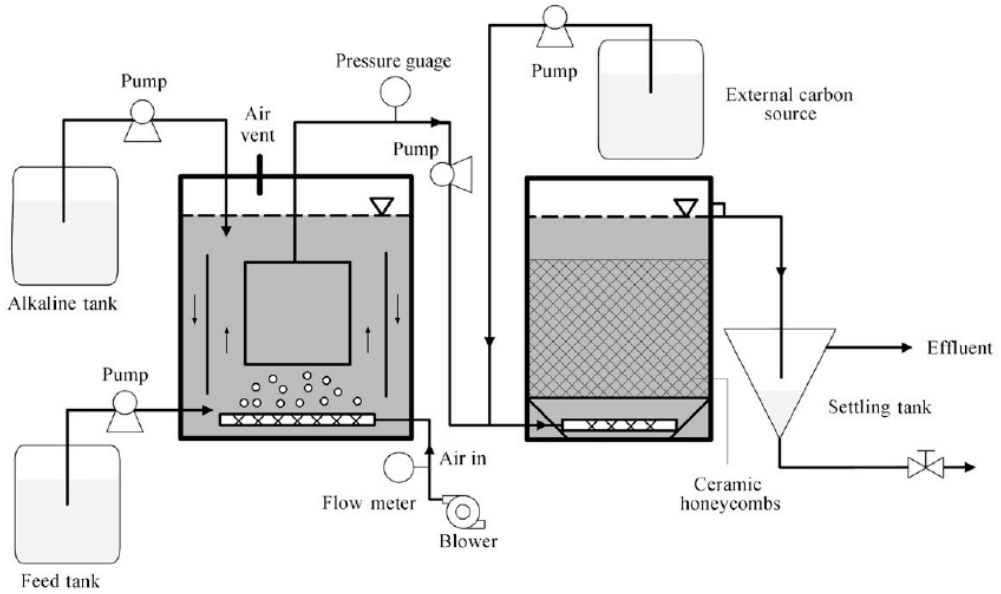


Figure 2.11. Combined Membrane Bioreactor and Anaerobic Packed-Bed Biofilm Reactor (Zhang, Zhou et al. 2009)

The configuration of this reactor appears to be less complicated than that of the membrane biofilm reactor. Unlike the sequencing batch moving bed membrane bioreactor, it maintains defined aerobic/anaerobic/anoxic zones that should provide better system control. The external carbon source could lead to residual BOD in the effluent, particularly after sloughing events. During the experiment the researchers maintained a pH of 7.8 through the use of an alkaline solution, which may or may not be necessary during practical applications. The various nitrogen species concentrations and removal efficiencies are illustrated in Figure 2.12.

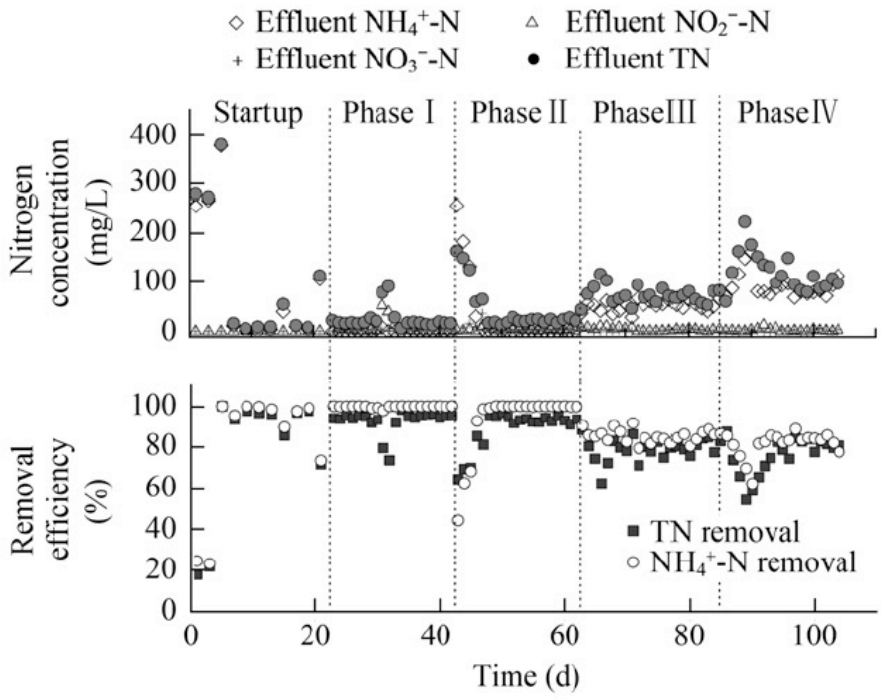


Figure 2.12. Variation of Nitrogen Concentration and Removal Efficiency in MBR-PBBR (Zhang, Zhou et al. 2009)

2.5.3. Systems that Separate Aerobic/Anoxic Zones Through the Use of Baffles or Hydraulics. These types of systems may provide less operational control than systems that completely separate zones and rely on pumps to return liquor, but their inherently simple design may outweigh the costs of consistent, maximum constituent removal. As such, these designs may appeal to decentralized wastewater treatment installations where capital cost and O&M costs are a major concern for the surrounding community.

The first of these examples is a reactor that is vertically stacked with the aerobic zone above the anoxic zone as shown in Figure 2.13 (Chae 2006).

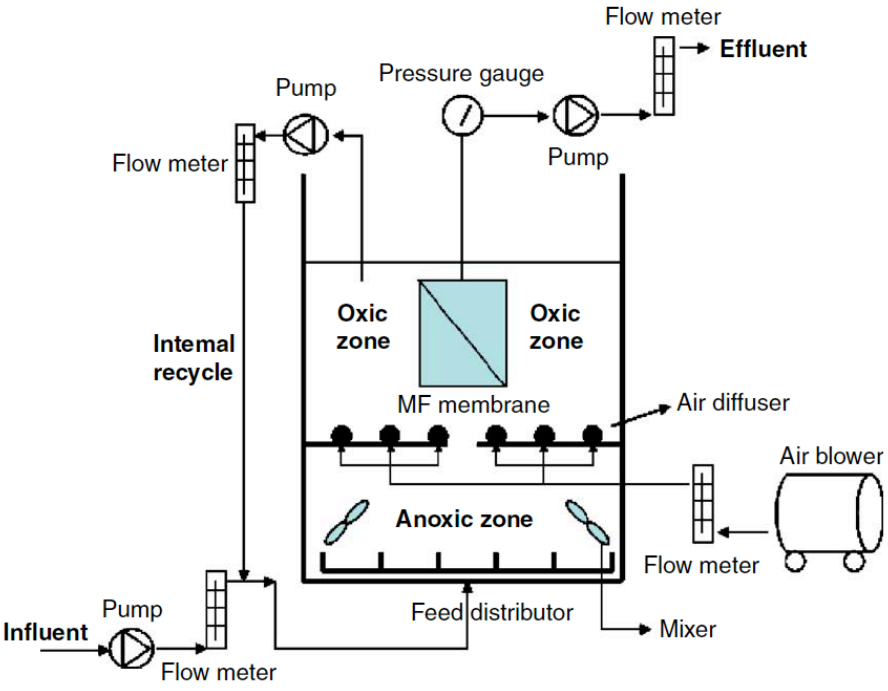


Figure 2.13. Vertical Submerged Membrane Bioreactor (Chae 2006)

Influent is fed to the anoxic zone, presumably eliminating the need for carbon addition. Mechanical mixers provide turbulence in the anoxic zone and force solids up toward diffusers, which create the oxic zone in the uppermost portion. Nitrified liquor is returned to the influent stream via mechanical pump, but this could possible be accomplished through the use of an airlift pump – eliminating the need for additional maintenance.

Results of the experiment can be seen in Figure 2.14 with nitrogen removal efficiency generally between 75 and 80%. The researchers focused on modifying the ratio of aerobic to anoxic zones, and separated their results based on the runs, which are listed in Table 2.4.

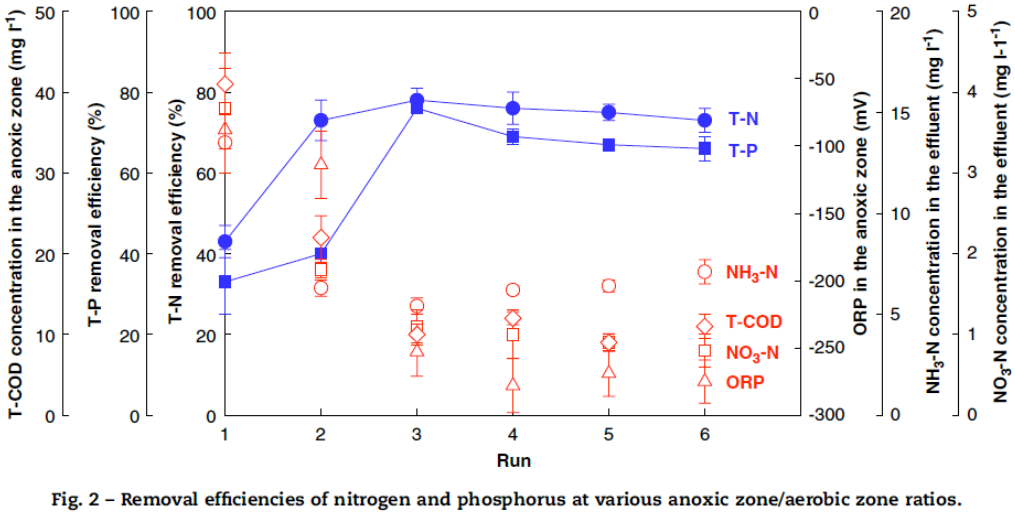


Fig. 2 – Removal efficiencies of nitrogen and phosphorus at various anoxic zone/aerobic zone ratios.

Figure 2.14. Removal Efficiencies of Nitrogen and Phosphorous at Various Anoxic/Aerobic Zone Ratios (Chae 2006)

Table 2.4. Anoxic/Aerobic Zone Ratios for Vertical Submerged Membrane Bioreactor (Chae 2006)

Condition	Run 1	Run 2	Run 3		Run 4	Run 5	Run 6
Reactor number	1	2	3		1	2	3
Effective volume of anoxic zone (A)	5.4 l	9.1 l	12.0 l		14.2 l	16.0 l	17.5 l
Effective volume of aerobic zone (B)	26.6 l	22.9 l	20.0 l		17.8 l	16.0 l	14.5 l
A/B ratio	0.2	0.4	0.6		0.8	1.0	1.2
SRT (days)				30			
HRT (h)				12			
Internal recycle ratio (%)				300			
Temperature (°C)	18.4±1.0	18.1±0.8	18.0±1.5		19.4±0.7	19.2±0.4	19.1±1.1
Operation period (days)	118	118	118		92	92	92

The second example of a simplified MBR using baffles to define zones comes from several experiments performed by Kimura et. al. (Kimura and Watanabe 2005; Kimura, Enomoto et al. 2007; Kimura, Nishisako et al. 2008). These experiments were all based on a single design as seen in Figure 2.15.

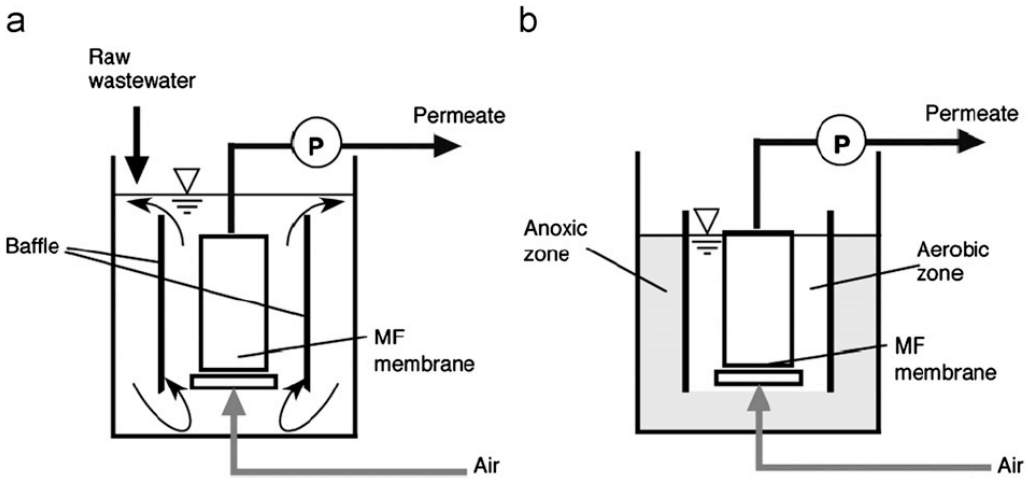


Figure 2.15. Baffled Membrane Bioreactor Operational Schematic (Kimura, Nishisako et al. 2008)

This unique design operates with liquor height determining the conditions in the reactor. When liquor height exceeds baffle height (as seen in “a” of Figure 2.15) the entire reactor is mixed via aeration from membrane diffusers. Once the liquor reaches a certain height the wastewater feed is paused and the reactor is allowed to drain. Once liquor height falls below baffle height (as seen in “b” of Figure 2.15) the outer zone become anoxic with the inner zone remaining aerobic. The researchers defined the cycle from fill through to the next fill as operation time per cycle (OTPC) with total aerobic to anoxic modes assumably being 1:1. The different OTPCs and their effect on total nitrogen concentration is illustrated in Figure 2.16, which came from the most recent reactor experiment and, assumably, the most refined reactor. Other studies by Kimura et. al. suggested other reactor iterations were not as effective.

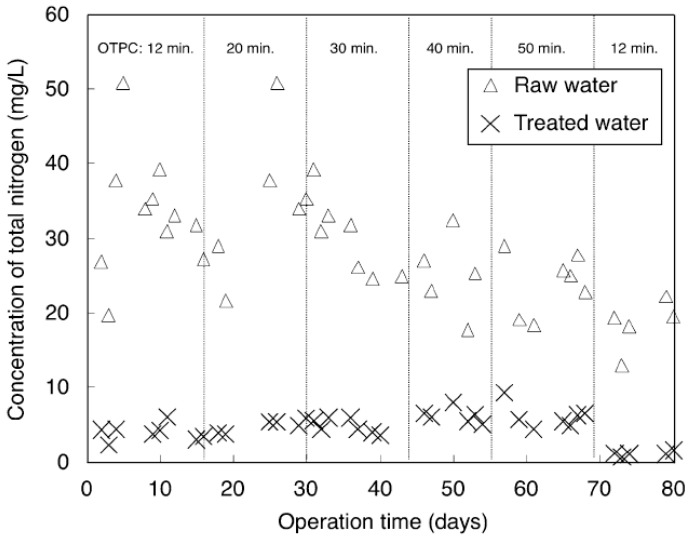


Figure 2.16. Influent and Effluent Total Nitrogen Concentrations (Kimura, Nishisako et al. 2008)

While this design is certainly unique in operation and configuration, it is questionable whether adequate mixing is provided to the anoxic portion of the reactor during the decant phase. Perhaps convective forces inside the aerobic zone are enough to mix the outer anoxic zone, but studies thus far fail to address that question. Furthermore, it is difficult to assess OTPC affect on total nitrogen removal with relatively short operational times between OTPCs and what appear to be constantly decreasing total nitrogen concentrations in the influent. However, this may be reflective of overall influent dilution that may have caused the amount of carbon available for denitrification to decrease as well.

3. GOALS AND OBJECTIVES

The goal of this research is to demonstrate the efficacy of a new MBR design for nitrogen removal evaluated by loading rate analysis (i.e., removal rate as a function of loading rate), while lowering the need for energy inputs when compared to conventional preanoxic systems. In order to attain this goal, the following objectives must be met:

- Design, build, and operate a conventional-type MBR with a single, aerobic volume to use as a control reactor.
- Design and construct a bMBR that will provide more effective nitrogen removal in terms of total nitrogen, ammonia-nitrogen, and nitrate-nitrogen while minimizing traditional energy input requirements (e.g., pumps).
- Operate both reactors over a similar period of time and under the same conditions (i.e., flow rates, temperature, influent characteristics, etc.).
- Compare reactors based on effective treatment of a range of typical wastewater constituents, with a focus on removal of various nitrogen species.
- Investigate difference of energy requirements for two reactors and estimate additional volumetric removal of nitrogen species as a function of the additional energy requirement.
- Provide overall assessment of reactor design based on effectiveness of treatment and energy use.

4. MATERIALS AND METHODS

4.1. OVERVIEW

This study consisted of two experiments. The first involved operation of a fully aerated MBR from late April 2009 until early January of 2010. The purpose of this first experiment was to provide baseline data for juxtaposition with data collected from the pilot-scale BBR. The second experiment allowed for more flexibility in reactor design, and a Baffled Membrane Bioreactor (bMBR) was operated from May of 2010 through late September of 2010. The focus of the second experiment was to maximize nitrogen removal through denitrification, since highly aerobic conditions in MBRs are inherently ideal for nitrification.

Samples from the first reactor (the MBR) were collected and analyzed for chemical oxygen demand (COD), total suspended solids (TSS), mixed liquor suspended solids (MLSS), and turbidity three times per week. Once a week a sample was analyzed for ammonia-nitrogen (NH_3-N), nitrate-nitrogen (NO_3-N), total nitrogen (TN), biological oxygen demand (BOD), dissolved oxygen (D.O.), temperature, and total phosphorous (TP). However, TP testing was halted in early August when phosphorous removal was no longer of interest for the Baffled Bioreactor (BBR). Samples from the second reactor (the bMBR) were taken and analyzed three times per week for COD, TSS, MLSS, D.O., temperature,

ammonia-nitrogen, nitrate-nitrogen, and total nitrogen due to the increased focus on nitrogen removal.

4.2. HOLLOW FIBER MEMBRANE MODULE

A hollow fiber membrane module was obtained from KOCH Membrane Systems to serve as the primary component of this experiment. This bench-scale module was based on KOCH's PURON line of membrane technologies. The module used in these experiments had an active surface area of 1.3 m^2 and external dimensions as noted in Figure 4.1.

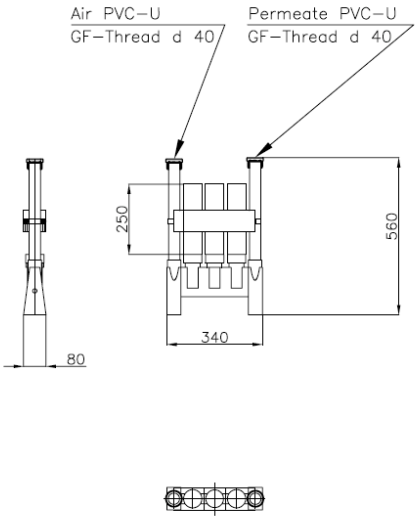


Figure 4.1. Hollow Fiber Membrane Module Diagram (KOCH Membrane Systems 2008)

The L1 braided hollow fibers for outside-in operation (KOCH specification on membrane type) had an outside diameter of 2.6 mm, a nominal pore size of 0.05 µm, and were composed of a proprietary polyethersulfone material. The unit was specified as having an operational temperature range from 5-40 °C, maximum filtration TMP of 9 psi, and a maximum backwash TMP of 14.5 psi. A picture of the unit during a cleaning operation can be seen in Figure 4.2.

Figure 4.2. Hollow Fiber Membrane Module Picture

Typical of membranes for wastewater solids separation, this module required regular cleaning. Occasional cold water scouring was necessary to remove severe caking that would randomly occur and cause filtration TMP to reach or exceed the maximum 9 psi. Chemical cleaning (a.k.a., recovery

cleaning) was also necessary. Every three to four months, based on filtration TMP, a recovery cleaning was performed per KOCH's instructions, which consisted of three-hour soaks in an alkaline, an acid, and a bleach solution. The alkaline soak was prepared by adding a 33% molar NAOH solution to tap water until a pH of 12 was reached. Adding a 30-32% molar HCl solution to tap water until a pH of 3 was reached created the acid soak. Finally, the bleach soak consisted of adding a NaOCl solution with 12% active chlorine by volume to tap water until a pH of 10.5-12 was reached.

4.3. REACTOR DESIGNS

Both reactors were fed wastewater that was diverted after the screens but before the grit chamber at the Rolla, MO Southeast Wastewater Treatment Facility. Once diverted, this stream flowed through a Mueller System Specialty 1/8"-pore basket strainer and to the reactor. Liquor levels in the reactors were controlled via float valve to maintain a consistent liquor level.

Permeate flowed into a control box that contained all mechanical and monitoring equipment. The basic design of the control box, shown in Figure 4.3, is illustrated by the system design shown in Figures 4.4 and 4.5. The illustrations show flow directions during permeate and backwash cycles, respectively.

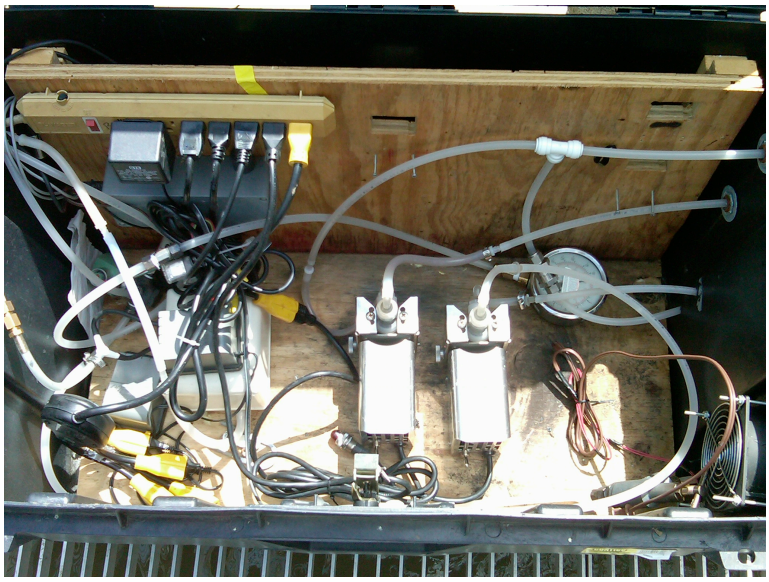


Figure 4.3. Picture of Reactor Control Box

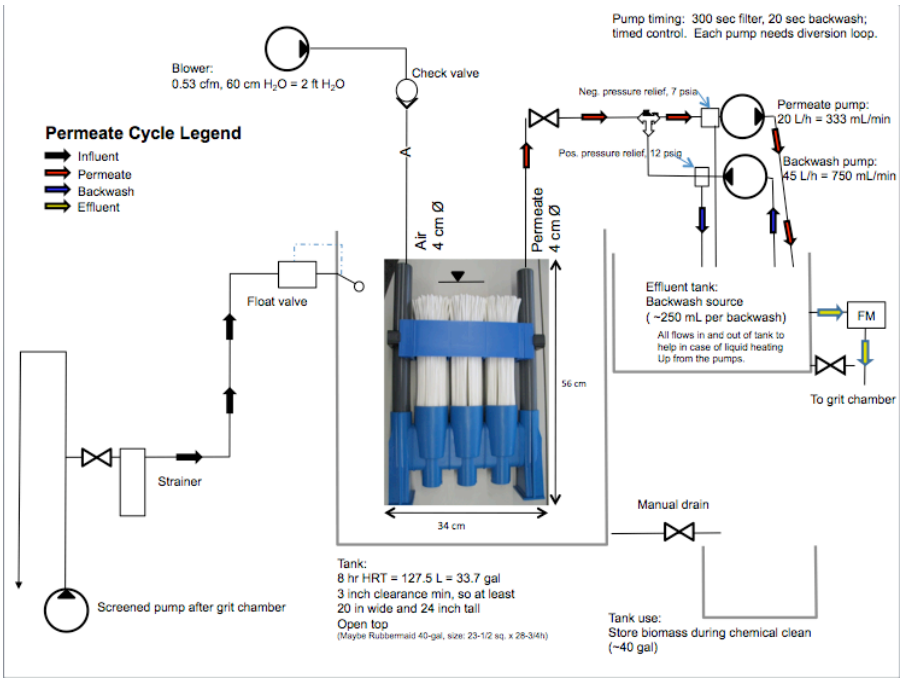
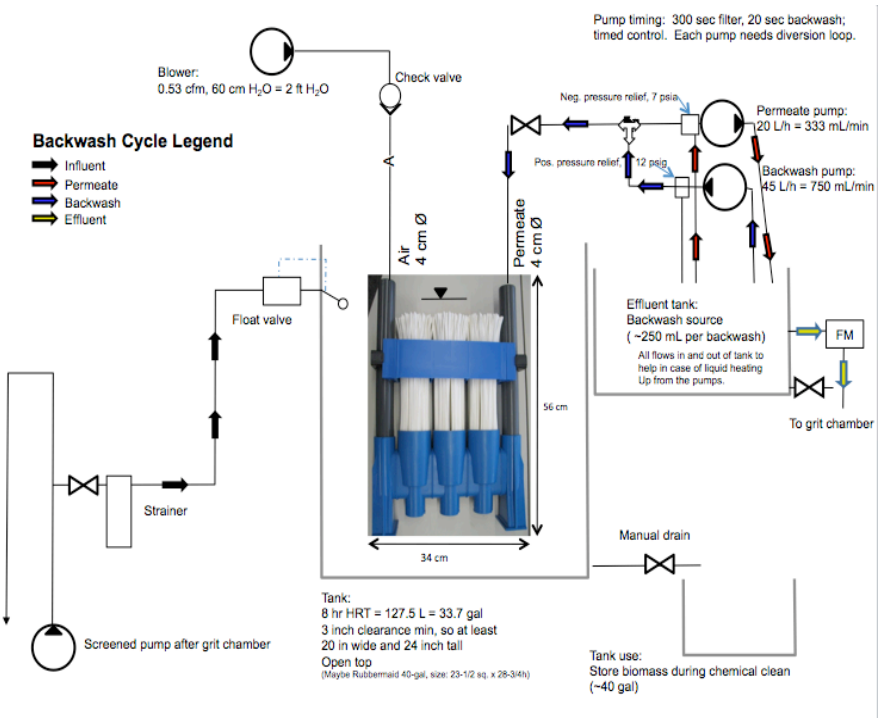


Figure 4.4. Schematic of System During Permeate Cycle



Figures 4.5. Schematic of System During Backwash Cycle

During the permeate cycle, water would enter the control box and flow through an ASCO Redhat II 3-way valve. This valve was used to control flow between permeate and backwash cycles. The electrical configuration that controlled this valve's timing was designed and installed by Brian Swift at Missouri University of Science and Technology.

Initially, once the permeate flow left the 3-way valve it would pass through a Microflow 100-1,000 mL/min flow meter (Blue-White Industries) before reaching the Model #K-07192-70 Cole-Palmer Bellows Metering Pump that carries a flow range of 173 to 1,730 mL/min. Soon after operation began it was realized that

these flow meters were not providing accurate data. It may be that paddle-type flow meters are not suitable for the variable flow produced by bellows pumps.

After flowing through the bellows pump the permeate would reach the effluent tank. The permeate was stored in this tank until being utilized as backwash during the backwash cycle or discharged as effluent once the liquor level in the tank reached the height of the discharge port. The first effluent tank was made of plastic and had the potential to allow light to enter the tank. A significant amount of algae was soon seen forming in both the effluent tank and the reactor tubing after the membrane. A metal tank was installed and the tubing wrapped in insulation due to concern that algae contamination in the tubing, effluent tank and membrane (transported during backwash cycle) might affect experimental results.

Transmembrane pressure was monitored by a WIKA 0-15 psi/-30-0 in-Hg pressure gauge. If the system was in permeate flow, the backwash was cycled back to the effluent tank through a positive pressure relief valve. When the system was in backwash cycle the permeate pump would create a vacuum and trip a negative pressure relief valve which allowed liquid in the effluent tank to be cycled through the permeate lines. As designed, the permeate and backwash ran consistently with the 3-way valve controlling the cycles and pressure relief valves preventing extreme pressures from being transferred to the membrane upon cycle change.

Pump failure occurred approximately three months into operation, and the electrical controls were modified. The new configuration allowed for the

backwash pump to be on only when the 3-way valve rotated to allow for the backwash cycle. This, along with rotating pumps between permeate and backwash duties, seemed to extend the life of the pumps and negated the need for the positive relief valve. Had another relay been added to the electrical controls the permeate pump could have operated in the same manner, further improving pump life and negating the need for the negative pressure relief valve.

Permeate and backwash flow rates were designed to be 333 mL/min and 750 mL/min, respectively. The 3-way valve was set to allow for 300 seconds of filtration followed by 20 seconds of backwash. Thus, total design flux was 1.29 L of permeate over the course of a permeate/backwash cycle, or 348 L/day. However, due to the lack of precise flowmeter readings we measured flow by stopwatch and graduated cylinder. Therefore, actual flow rates were maintained at 320 mL/min and 720 mL/min for permeate and backwash, respectively. This resulted in actual total flux per cycle of 1.36 L, or 367 L/day.

4.3.1 Design and Operation of the MBR. The MBR that operated April through December of 2009 consisted of the KOCH membrane in a 35-gallon container with a 98 L operational volume as seen in Figure 4.6. Air to the membrane module was provided via an ALITA AL-15A air pump rated for 15 L/min. However, the maximum measured airflow rate onsite was never greater than 7.9 L/min. The airflow was measured in situ by placing a TSI 4000 Series air flow meter in the airline.

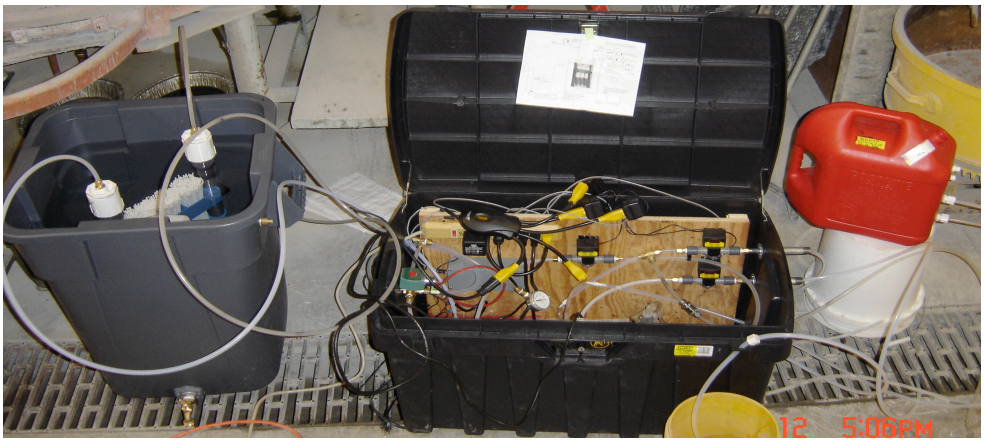


Figure 4.6. MBR in 35-Gallon Container with Control Box and Backwash Tank

The primary challenge of operating this reactor came with the extremes in season temperature. Small flow rates and direct exposure of the reactor and control box to the elements resulted in operational temperatures that closely related to ambient temperatures. Summer in central Missouri saw daily highs exceeding 32°C on more than one occasion. An exhaust fan was added to the control box for increased ventilation and tubing inside the control box were insulated to help reduce condensation.

Winter brought temperatures that frequently dropped below -5°C. At one point ice exceeding 1" in thickness formed on top of the liquor in the reactor. Heat tape was used to prevent the effluent and influent lines from freezing, and a cattle deicer was inserted into the reactor to help keep temperatures above freezing. These low temperatures seemed to have negative affects on pump operation, and maintaining consistent flow rates proved to be extremely difficult if not impossible.

4.3.2 Design and Operation of the bMBR. The bMBR, operated May through September of 2010, was designed for advanced nitrogen removal by creating separate anoxic and aerobic zones within the reactor. While the first MBR seemed to provide some nitrification and denitrification, these processes were unreliable. There was no way to control anoxic and aerobic volumes within a combined reactor. The bMBR gave us the ability to control, monitor and maintain the necessary environmental conditions to encourage specific microbial functions. The separate anoxic and aerobic zones, as seen in Figure 4.7, had operational volumes of 74 L and 33 L, respectively. The reactor, control box and effluent tank were placed in operation at the plant as shown in Figure 4.8.

Figure 4.7. bMBR with Anoxic Zone and Aerobic Zone

Figure 4.8. bMBR in Operation

The bMBR utilized the same control box and components as the first, aside from the blower. Initially, the ALITA air pump was used for coarse-bubble membrane scouring. However, the second reactor seemed to retain significantly higher concentrations of biomass that resulted in severe caking of the membrane module and D.O. concentrations below the 2 mg/L commonly recommended for complete nitrification. A fine-bubble diffuser was installed to help alleviate low D.O. concentration, followed by air stones that allowed for fine bubble diffusion to be more evenly dispersed in the reactor. However, these actions did not seem to remedy the problem.

Soon thereafter a Danner MFG Inc. AP-20 air pump was installed in place of the ATILA air pump. This occurred once operation recommenced after the first recovery cleaning of the membrane on July 15. Although the Danner had a listed flow rate of 27.9 L/min (1,700 in^3/min), the airflow rate measured onsite was

never greater than 15.5 L/min. The additional volume of coarse-bubble air seemed to increase D.O. and provide better scouring, and the ALITA was used to provide fine-bubble diffusion to increase D.O. concentration in the reactor. Airflow from the ALITA was adjusted daily via a bleed-off valve to control the D.O. concentration.

A Robbins and Myers 1/8 HP 1,675 RPM motor provided mixing for the anoxic zone. Influent would enter the reactor through the float valve into the anoxic zone. The anoxic mixer also produced hydraulic lift using two propellers on a single shaft, which allowed some of the anoxic liquor to flow over a weir and into the aerobic zone with the membrane module. A profile of the anoxic/aerobic weir and a top-down schematic of the reactor can be viewed in Figure 4.9.

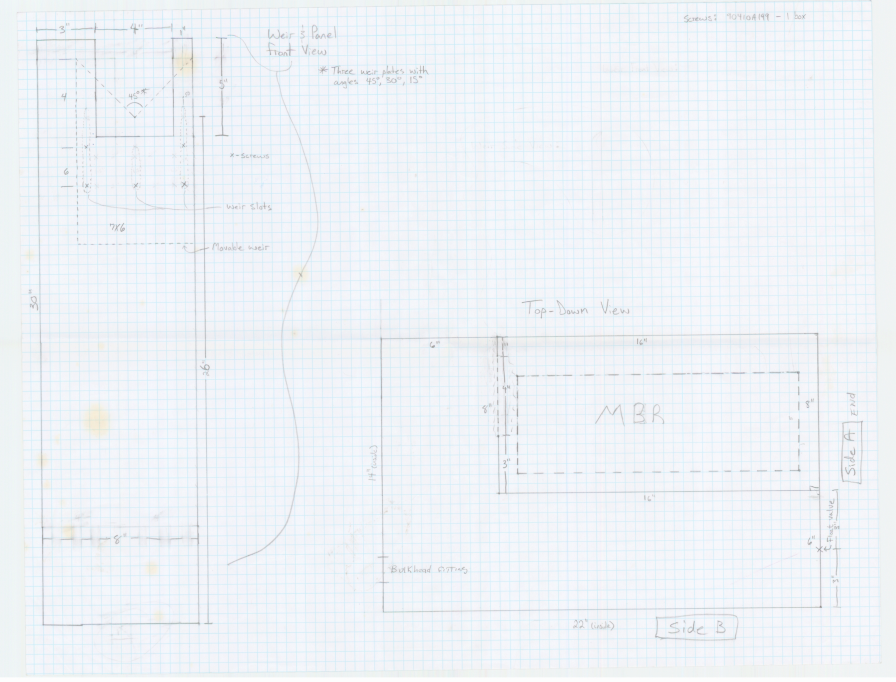


Figure 4.9. Anoxic/Aerobic Weir and Top-Down Reactor Schematic

Figure 4.10 shows a profile of the reactor with the anoxic zone on the left, aerobic on the right, and the aerobic/anoxic baffle separating the zones.

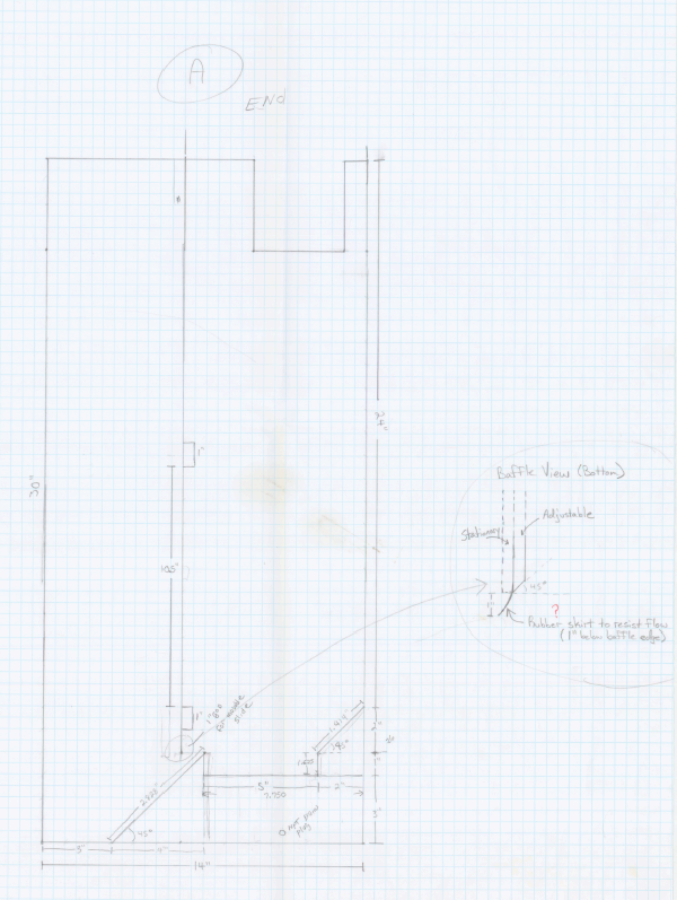


Figure 4.10. Profile of Reactor with Anoxic Zone, Aerobic Zone, and Aerobic/Anoxic Baffle

Once in the aerobic zone the mixed liquor would either be drawn toward the membrane where liquid/solid separation would take place, or it would flow under a baffle and back into the anoxic zone. Constant flow under the aerobic/anoxic baffle was achieved by increasing the hydraulic head in the

aerobic zone relative to the anoxic zone, thanks to the hydraulic lift induced by the anoxic mixer. Flow rate to the aerobic zone and return to the anoxic zone were controlled by adjusting the height of the weir and depth of the baffle. Note placement of the membrane module. The diffusers of the module were countersunk into the reactor and surrounding horizontal surfaces were sloped toward the anoxic zone at a 45° angle to encourage solids migration.

Figure 4.11 shows a profile of the bMBR and illustrates flow patterns within the reactor.

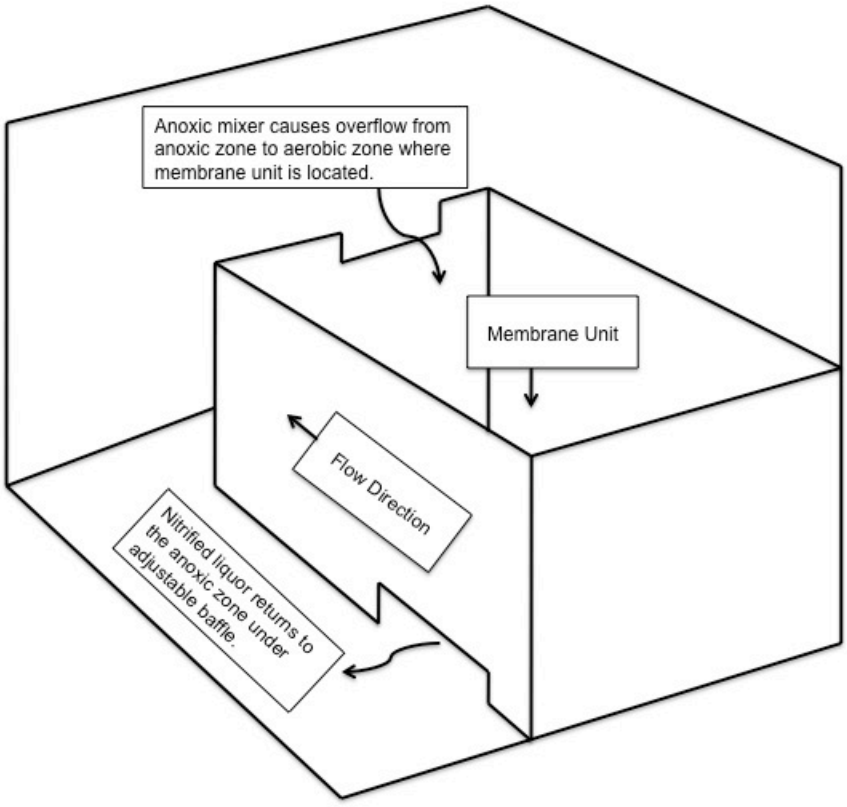


Figure 4.11. Conceptual Schematic of the bMBR with Flow Direction

The outer volume receives influent to the reactor and was maintained under anoxic conditions. The inner volume was aerobic and contained the membrane unit. To place Figure 4.11 in proper perspective, Figure 4.7 should be referenced to see the position of the float valve, anoxic mixer and membrane unit.

4.4. MEASUREMENTS

Samples were taken onsite with Nalgene containers with volumes of 250 mL and 1 L, as well as 50-mL Falcon vials. Sludge characteristics were evaluated onsite using a makeshift sludge judge with tubing attached to 1/2" PVC with marks for distance/depth. A TSI 4000 Series air flow meter was used to measure airflow rates with a range of 0-300 L/min and an accuracy of 2% of reading.

BOD was measured using HACH Formula 2533 nitrification inhibitor and an YSI 5000 with an YSI 5010 D.O. probe. Incubation occurred in a Fisher Scientific Low-Temperature Incubator. D.O. meter calibration was completed by placing the D.O. probe in a 250-mL BOD bottle with approximately 1″ of DI water at the bottom. The meter was given time to stabilize (usually no more than five minutes) and the resulting concentration saved as the saturation point. The range for this unit is 0-60 mg/L with an accuracy of ± 0.1% plus 1 least significant digit.

BOD results during operation of the MBR may not be valid. Standard Methods stipulates a blank, seed determination and glucose-glutamic acid standard should be run every time a BOD analysis is performed, and if any basic requirements of these three are not met at the end of the 5-day incubation period than the sample is invalid (Loftus 2003). Moreover, pH, temperature and ending D.O. were not necessarily considered at the beginning and ending of the analysis. Since the glucose-glutamic acid standard under ideal conditions should produce a range of ± 30.5 mg/L, that's the range chosen for the data. Not running duplicates and/or at least three dilutions of the same sample makes calculating range extremely difficult if not impossible.

Total suspended solids were measured using Whatman 934-AH Glass Microfiber Filters (Cat: 827047), a Denver Instrument ST-234 digital balance, a KNF Labs filtration pump, and a Fisher Scientific Isotemp Oven. If volatile suspended solids were measured an American Scientific Products FP-31 oven would be used during final incubation. Suspended solids have a method detection limit (MDL) of 10 mg/L when samples of 100 mL are poured through the vacuum filer (EPA 1999). However, this MDL may change depending on the volume of sample used and appears to be determined based on margin of error. For example, margin of error when analyzing a 100-mL sample is 10 mg/L because of the "10" multiplier to convert to mg/L. Therefore, the MDL and range of a 10-mL sample would be 100 mg/L due to the "100" multiplier (Loftus 2003). Also, it should be noted that glass filters were not washed, dried and weighed to verify a constant weight before analyzing samples, as required by Standard

Methods 2540D or EPA method 160.2. Therefore, some positive error may be present in the data (i.e., TSS, VSS, MLSS and MLVSS concentrations may be listed to be greater than they actually are).

pH was monitored using an ORION Perphect LogR Meter Model 370, and turbidity was measured using a HACH 2100P portable turbidimeter. The meter was calibrated using a two-point calibration method with standards of pH 7 and 10. This unit has a range of 0-14 pH and an accuracy of ± 0.005 at 25°C.

D.O. and temperature were measured in-situ using a YSI model 58 D.O. probe and meter, which has a range of 0-20 mg/L, a resolution of 0.01 mg/L and an accuracy of 0.03 mg/L. Calibration for the D.O. meter was performed in the same manner as calibration for the D.O. probe when testing BOD.

Turbidity of the MBR's influent and effluent were measured using a HACH 2100P portable turbidimeter with HACH standards for calibration. This instrument has a range of 0.01 to 1,000 NTU with an accuracy of ± 2%.

Many of the constituents were measured using HACH chemical reagents. High-range COD was measured using HACH HR Digestion Solution Cat: 21259-15 reagent (20-1,500 mg/L), incubated in a HACH DRB200 and analyzed using a HACH DR2010 spectrophotometer. This method has a MDL of 5 mg/L and a range of ± 18 mg/L. Calibration for the high-range COD test was performed by placing a sample volume of DI water into a test reagent, incubated with samples, and used to zero the spectrophotometer. Calibration standards were discarded after five days of use.

Low-range COD was measured using HACH LR TNTplus 821 (3-150 mg/L, ± 3 mg/L), nitrate-nitrogen with HACH LR TNTplus 835 (0.23-13.5 mg/L NO_3^-, ± 0.63 mg/L), phosphate-phosphorous with HACH LR TNTplus 844 (1.5-15.0 mg/L PO_4^{3-}, ± 0.5 mg/L), and total nitrogen with HACH TNTplus 827 (5-40 mg/L, ± 1 mg/L). All were analyzed using a HACH DR2800 spectrophotometer. If incubation was necessary a HACH DRB200 was used.

Ammonia-nitrogen was initially measured using an Orion ammonia electrode (Orion 95-12) with a range of 0.01-17,000 mg/L. This was done by adding 25 mL of sample to a 50-mL Falcon vial. Calibration would be performed using two standard solutions before adding 0.5-mL pH-adjusting ISA to the sample vial and taking the reading. However, analysis using this probe seemed to become unreliable in early 2009 and we quickly switched to HACH methods using reagents TNT plus 832 for high-range (2-47 mg/L, ± 1 mg/L) and TNT plus 831 for low-range (1-12 mg/L, ± 0.4 mg/L). The overall effects on data due to the initial use of the probe versus HACH reagent are assumed to be negligible do to the minimal amount of time the probe was used.

Flow rate over the anoxic/aerobic weir in the bMBR was targeted to be 999 mL/min which, when the permeate flow was kept at 333 mL/min, would induce a recycle rate of 666 mL/min – 200% of inflow rate. This flow rate was determined noting the height of the flow above the crest of the weir. Equation 5 was used to calculate the necessary height (LMNO Engineering 2007).

$$Q = 4.28C * \tan\left(\frac{\theta}{2}\right) * (h + k)^{5/2} \quad (5)$$

Where Q = flow rate (cfs)
h = height over weir (ft)
$k = 0.01449 - 0.000339\theta + 3.298x10^{-6} * \theta^2$
$C = 0.607 - 0.000874\theta + 6.104x10^{-6} * \theta^2$
θ = notch angle in degrees

Statistical analysis was performed assuming standard normal distribution of sample data within the population, and utilizing Student's *t* distribution as developed by William S. Gosset (Berenson 1999). To determine confidence intervals for sample data with 95% confidence where population mean and standard deviation is unknown, Equation 6 was used.

$$\overline{X} \pm t_{n-1}\frac{S}{\sqrt{n}} \quad (6)$$

Where $\overline{X}$ = Sample mean

t_{n-1} = Critical value of the *t* distribution (determined in Excel)

S = Sample standard deviation

n = Number of samples

Energy use of the anoxic mixer was measured after experimentation using an IDEAL Clamp-Pro 600 Amp Clamp Meter (Catalog #61-744), with a basic accuracy of ± 1.7%. To gauge current the mixer was placed in a 5-gallon bucket

filled with tap water. The mixer required 2 amps during operation, which translates into 230 watts of power when multiplied by the 115-volt mixer rating.

4.5 REACTOR OPTIMIZATIONS

Experimental procedures varied from reactor to reactor. Little was done to the first reactor to improve performance. It was the control treatment system in a larger experiment, and constant performance, rather than optimum performance, was desired.

The second reactor saw more flexibility in terms of system optimization. However, because the experiments were conducted at an operating wastewater treatment plant using municipal wastewater, it was difficult to isolate variables. For example, on occasion D.O. was extraordinarily high and denitrification suffered as a result. Simply lowering airflow to the reactor might remedy the problem for a few hours, but the influent constituent concentrations would change due to diurnal variation or I&I effects – thus making precise operation difficult if not impossible. Had the experiment been run under laboratory conditions or monitored more frequently, better control over D.O. concentrations in the different zones would have made treatment more effective. Table 4.1 displays key dates, observations, and/or modifications made to the reactor.

Table 4.1. Dates, Observations, and Modifications to bMBR

Date	Observations	Modifications
5/20/2011	Initiate operation	
5/24/2011	One of three coarse-bubble diffusers clogged	Attempted to shake unit to clear, but seemingly to no avail
5/25/2011	Diffuser unclogged by itself	
6/1/2011	Two of three coarse-bubble diffusers clogged. Negative pressure during permeate cycle was 15”-Hg before scour and 5”-Hg after scour.	Removed membrane from liquor and scoured with tap water to remove caking and clear diffusers.
6/4/2011	D.O. seems to be consistently low. TMP increased to near 15”-Hg. May have to scour every few days to maintain operation.	Added fine-bubble diffuser and scoured membrane.
6/7/2011	Flowrate over the anoxic/aerobic weir seems to be greater than 999 mL/min.	Lowered aerobic/anoxic baffle one notch (1/4”) to encourage greater detention time in aerobic zone.
6/8/2011	TMP too high, but flow over weir looks to be closer to 999 mL/min based on flow height over weir.	Scoured membrane
6/9/2011		Decreased permeate cycle duration to help lower rate of TMP increase.
6/14/2011		Scoured membrane
6/22/2011		Scoured membrane
6/23/2011		Scoured membrane and switched fine-bubble diffuser to air stones to help disperse fine bubbles.
6/25/2011		Scoured membrane
6/28/2011	Effluent bypass left open on Friday (6/25) resulting in backwash tank running dry. Therefore, no backwash over weekend.	
6/30/2011	Two of three coarse-bubble diffusers clogged	Scoured membrane
7/6/2011		Scoured membrane
7/7/2011		Scoured membrane
7/9/2011		Scoured membrane
7/11/2011	One of three coarse-bubble diffusers clogged	Scoured membrane
7/14/2011		Began recovery cleaning of membrane with chemicals

Table 4.1. Dates, Observations, and Modifications to bMBR (Continued)

Date	Observations	Modifications
7/15/2011		Operation resumed after cleaning. New blower installed for coarse-bubble (14.4 L/min) and old coarse-bubble blower switched to fine-bubble diffuser (5.9 L/min).
7/19/2011	MLSS appeared to be very dilute, possibly due to significant amount of rain over past few days. D.O. greater than 5 mg/L.	
7/23/2011	D.O. had been consistently high, possibly affecting denitrification.	Decreased airflow to fine-bubble diffuser.
7/26/2011	D.O. too low	Increased airflow to fine-bubble diffuser.
7/29/2011	Sludge judge appeared to show scattered areas of high-density sludge.	Raised aerobic/anoxic baffled one notch (1/4”) to increase mixing.
7/30/2011		Lowered aerobic/anoxic baffle one notch (1/4”) as raising baffle had no apparent affect on mixing. Reduced airflow to fine-bubble diffuser.
8/4/2011	Flow from anoxic to aerobic over weir seemed inconsistent, with backwash back into anoxic zone.	Moved weir as high as possible and dropped baffled one notch.
8/6/2011	Solids in reactor almost completely settled. Liquor at reactor surface clear. D.O. 5 mg/L above sludge blanket and 0.2 mg/L in sludge blanket.	Scoured reactor, which also seemed to disperse sludge blanket.
8/23/2011		Scoured membrane. Raised baffle ½ notch.
8/25/2011	Arrived at plant to find bMBR near dry. Lab partner switched off grinder pump at breaker box.	Filled reactor with wastewater and restarted.
8/29/2011		Lowered float valve to decrease liquor height in anoxic zone and raised baffled ½ notch.
8/30/2011		Lowered weir
9/8/2011	Two of three coarse-bubble diffusers clogged.	Scoured membrane
9/24/2011		Operation ceased

Experimental procedures consisted of testing influent, effluent, and reactor parameters at least three times per week. The reactor was observed almost everyday. D.O. concentrations in the aerobic and anoxic zones, mixing observations, and flow rate over the anoxic/aerobic weir would influence modifications to anoxic/aerobic weir height, aerobic/anoxic baffle depth, and airflow to fine bubble diffusers. The reactor would then be allowed to reach steady state and observed over several days.

5. RESULTS AND DISCUSSION

5.1 MBR DATA

The MBR was operated as the control treatment for a separate experiment from April through December of 2009. The KOCH membrane cartridge was placed in a reactor with an operating volume of 103 L. There were no defined boundaries between anoxic and aerobic zones, and whatever nitrification and denitrification occurred were presumably due to either pockets of high and low D.O. in the reactor or by limited D.O. diffusion into flocs.

Grab samples from the influent, effluent and mixed liquor were taken for testing purposes. Temperature and D.O. (Figures 5.1 and 5.2, respectively) in the reactor were checked in-situ roughly once a week. D.O. and temperature appear to have an inverse relationship, as expected. The D.O. in Figure 5.2 was taken at or near the surface of the reactor. Positional sampling suggested D.O. concentrations throughout the reactor were relatively constant. However, measurement of D.O. below the MBR diffusers revealed a D.O. concentration of essentially 0 mg/L. It is unknown whether this contributed to denitrification. The conditions seem to make it possible, but the volume below the diffusers is assumed to have been unmixed suggesting little mass transport out of this zone.

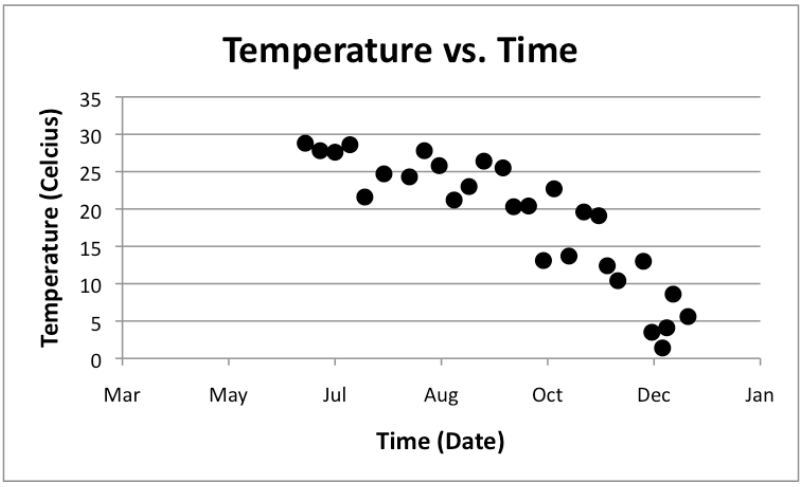


Figure 5.1. MBR Temperature

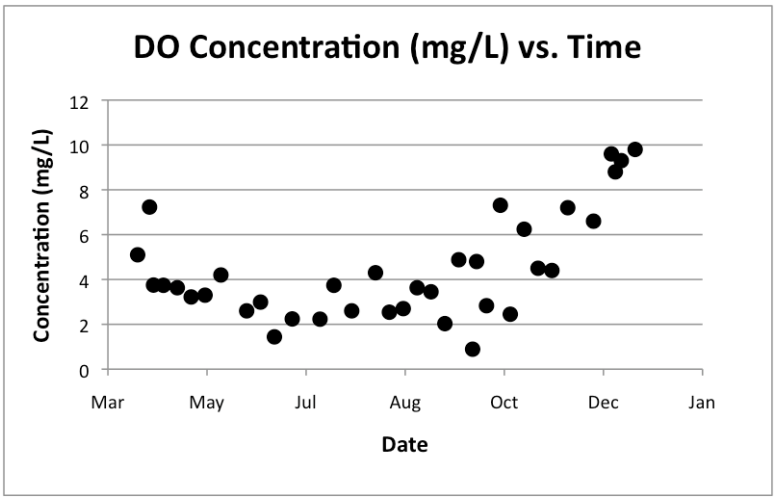


Figure 5.2. MBR Dissolved Oxygen

The great variation of influent TSS concentration (observable in Figure 5.3) may be attributable to rain events and natural variations in sewage, as well as any bias introduced by using grab samples. Due to sample size analyzed,

error bars for individual samples were ± 50 mg/L for influent and ± 1 mg/L for effluent. The mean TSS concentration (Figure 5.3) in the effluent was 0.79 ± 0.2 mg/L with 95% confidence.

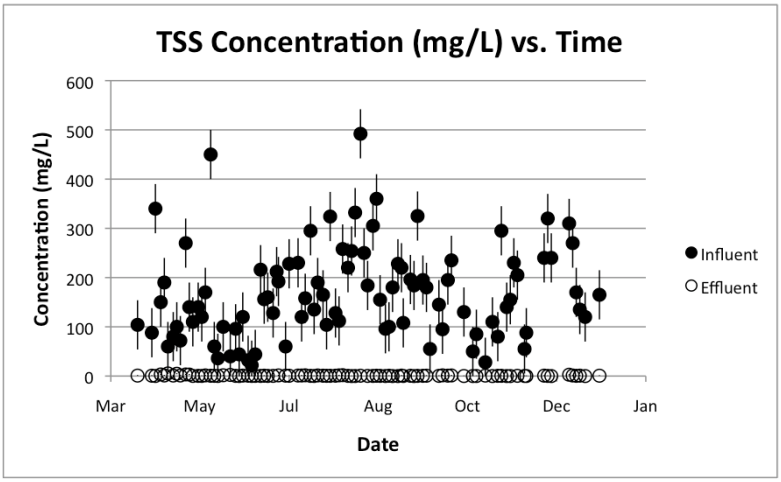


Figure 5.3. MBR Total Suspended Solids

The MBR seemed to have a dense stratified layer that formed roughly around the height of the diffusers, about three inches from the bottom of the reactor, with a higher MLSS concentration than the mixed liquor above the diffusers (Figure 5.4). This layer, which did not display characteristics of aerobic sludge as previously noted, was much more concentrated than the rest of the reactor. Individual error bars in Figure 5.4 were determined to be ± 1,000 mg/L for the bottom layer and ± 50 mg/L for the surface layer. The bottom layer had a

mean concentration of 1,700 ± 380 mg/L while the surface had a mean MLSS concentration of 39.7 ± 12.4 mg/L.

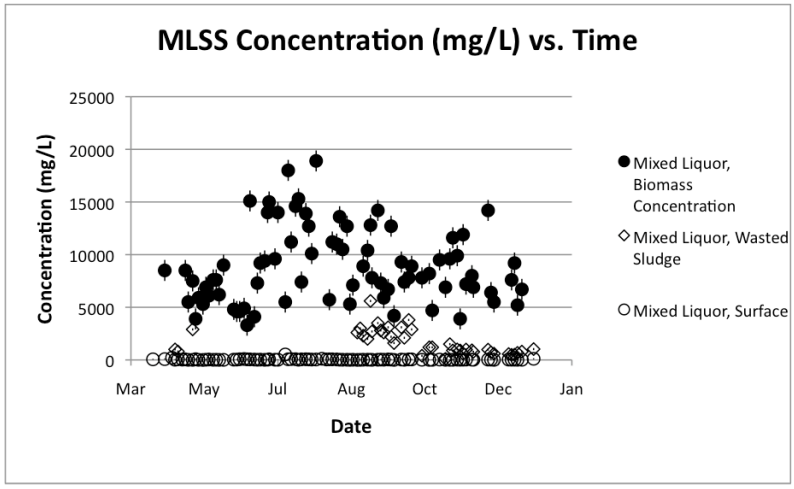


Figure 5.4. MBR Mixed Liquor Suspended Solids

When sludge was wasted, the membrane unit was rocked to help mix the reactor and scour the membrane. The wasted sludge biomass concentration may better represent the concentration of biomass throughout the entire reactor than samples taken in the upper portion of the tank or from the bottom. Some discrepancy may exist due to concentrated sludge passing through the wasting valve before complete mixing had taken place, particularly if solids or flocs were too dense to readily suspend. This discrepancy is represented in the large confidence interval for the mean concentration. The error bar for the wasted sludge was determined to be ± 100 mg/L. The mean concentration of MLSS in

the wasted sludge was 8,800 ± 790 mg/L. However, much of this biomass may have been rendered essentially inactive due to minimal amount of mixing at the bottom of the reactor.

High removal rates of total nitrogen and COD may have been contributed more to liquid/solid separation by the membrane than biological degradation, given the relatively dilute concentration of well-mixed biomass in the reactor. To better determine this in the future it may be advisable to determine soluble COD and total nitrogen by passing a sample through a filter with nominal pore size roughly equivalent to those of the membrane before performing analysis.

Prior to mid-September, sludge was rarely wasted from the MBR. The stratified nature of this biomass is reflected in Figure 5.4; the surface samples had a low biomass concentration, while the little-mixed sludge below the diffusers was quite concentrated. Once sludge wasting started, the concentration of the mixed, wasted sludge was measured. We noted on 2 September that measurements of what was thought to be anaerobic sludge at the reactor's bottom were becoming inconsistent. Apparently, enough of this layer had been wasted that it was present in clumps at the bottom of the reactor rather than in a defined layer as before. The mixed liquor at the reactor's surface remained relatively dilute.

Pump failure occurred in September 25 due to motor wear. The result was a stop in flow for two days. While waiting on a replacement bellows pump a peristaltic pump was used for permeate flow. However, pressure less than -3" Hg

caused the soft tubing feeding the peristaltic pump to collapse and greatly reduce flow. Normal operation occurred after about a week of reduced flow.

The concentration of solids in the wasted sludge decreased after this pump failure and with the onset of cooler temperatures. Most solids began to resemble flakes, and it may be that flocs greatly increased in size due to the temperature decrease. Also, sampling of the biomass concentration became increasing difficult over the same time period. Sludge levels at the bottom of the reactor became inconsistent, resulting in data without a discernable trend.

Turbidity (Figure 5.5) is another method of characterizing suspended solids – particularly when the fluid contains a very dilute concentration of solids.

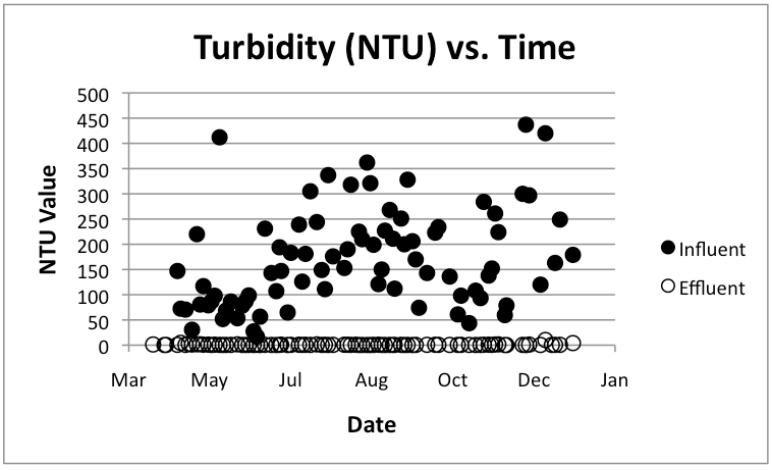


Figure 5.5. MBR Turbidity

The mean turbidity of MBR effluent was 0.92 ± 0.26 NTU. Influent pattern of TSS concentration and NTU seems to be consistent with each other,

suggesting a strong relationship. As a point of reference, the allowable standard for drinking water in the United States is 0.3 NTU, with many utilities striving for a goal of 0.1 NTU.

The effluent of the MBR contained organic solids, as measured by VSS as shown in Figure 5.6, which were on average 82% of the TSS, but could have been anywhere between 45% to 100% of the TSS. The error bars for VSS were assumed to be the same as TSS (Figure 5.3).

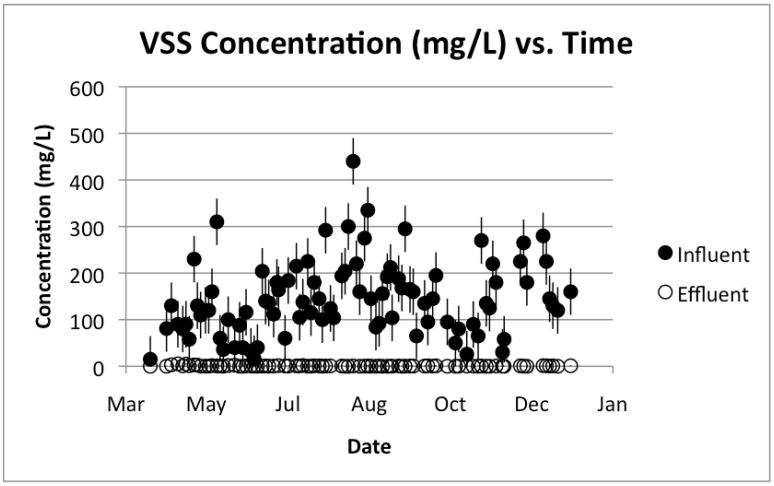


Figure 5.6. MBR Volatile Suspended Solids

It is quite possible that some of the organic material measured in effluent VSS was algal growth that was sloughed from the permeate lines. Evidence of this was noticeable as a green coloration on the filter after filtration. This discoloration of the filter decreased steadily after the tubing was wrapped with

insulation in early July. A faint tinge was periodically observed until after variable flowrate began to occur in December, presumably due to low temperature effects on the pumps. At that point the pump seemed to skip pulses of liquid and overall flowrate was unknown, and effluent displayed a rust-red coloration for 7-10 days. The color then decreased in intensity and was not noticeable at the end of the testing period in early January. Average effluent VSS concentration was 0.65 ± 0.2 mg/L.

During the first several months of operation, no sludge wasting occurred. During this time the endogenous decay of the biomass was presumably equal to the growth rate, allowing for an infinite biomass age. As a result, nonorganic particles and non-biodegradable organics were believed to be building up in the reactor, which led to the decision to begin wasting sludge on July 24th. If a trend line is applied to the mixed liquor biomass concentration shown in Figure 5.7, it appears that the concentration remained fairly steady through mid-June when the temperature began to rise and flows into the plant decreased. The decrease in flows due to dry weather could have resulted in an increase in BOD. Because the flow rate into the MBR was constant, the increased BOD concentration may have resulted in a marked increase in MLVSS concentration until sludge wasting commenced. Error bars for MLVSS were assumed to be the same as MLSS (Figure 5.4). Mean concentrations of MLVSS at the reactor's surface, below the diffusers when collected by sludge judge, and during sludge wasting were 33 ± 9.2 mg/L, 1,380 ± 310 mg/L, and 6,500 ± 530 mg/L, respectively.

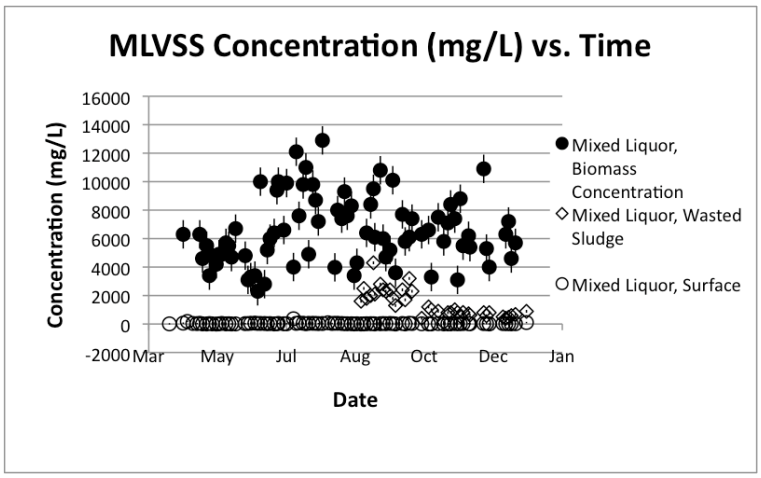


Figure 5.7. MBR Mixed Liquor Volatile Suspended Solids

COD concentrations in the MBR are shown in Figure 5.8. COD removal in the MBR was excellent throughout the testing period. COD concentration in the effluent averaged 14.9 ± 1.72 mg/L.

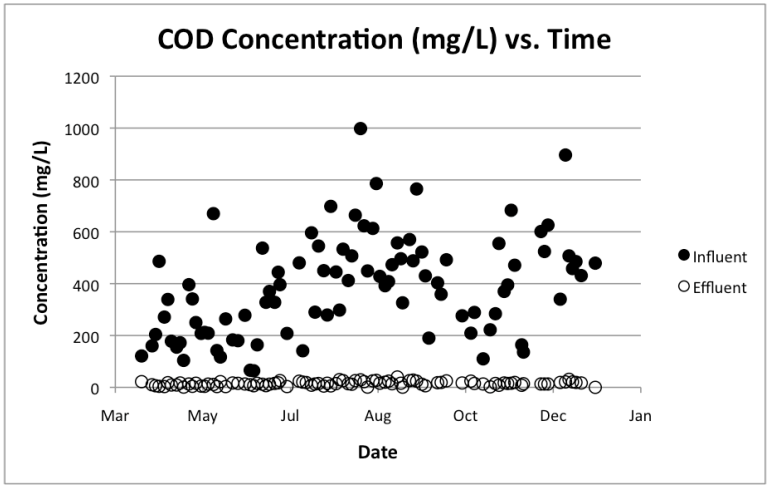


Figure 5.8. MBR Chemical Oxygen Demand

As previously suggested, the concentrations of BOD, TN, and TSS in the influent increased as rain events became less frequent and sewer flows presumably became less dilute.

Figure 5.9 shows that the effluent BOD of the MBR averaged 1.71 ± 0.55 mg/L – well below discharge limits set by the Ten State Standards and federal regulations. However, all BOD results are considered invalid under Standard Methods due to reasons listed in Section 3.3.

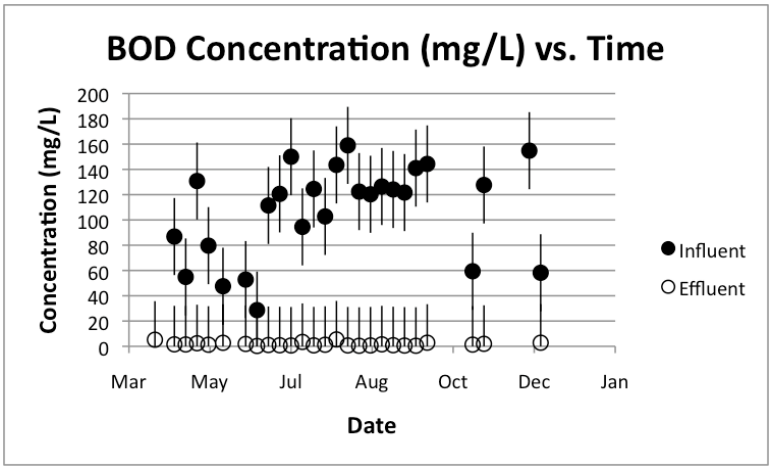


Figure 5.9. MBR Biological Oxygen Demand

Total nitrogen (TN) removal (Figure 5.10) in the MBR varied. Most likely much of the variance in removal efficiency can be attributed to lower operating temperatures and MLSS/MLVSS concentrations, but with so many variations in operating conditions it is difficult to draw solid conclusions.

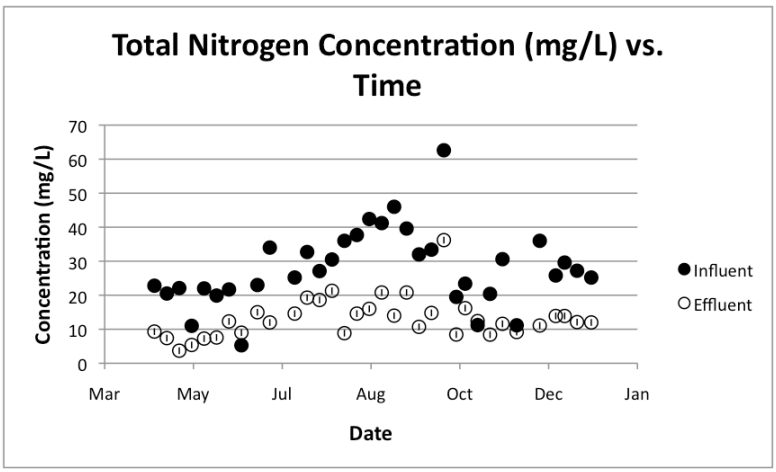


Figure 5.10. MBR Total Nitrogen

It should be noted that instances when the influent TN was less that the effluent TN of the MBR (Figure 5.10) occurred just after heavy rain when the inflow to the plant had become dilute due to inflow and infiltration. The residence time of the reactor is such that effluent samples reflect the influent characteristics from roughly six hours before; thus the effluent sample was taken before the reduced TN concentration had passed through the reactor. While it initially appeared nitrogen removal increased after sludge wasting began on July 24th, statistical analysis suggests that it is possible the difference in concentrations may be attributable to natural variations. The mean total nitrogen concentration in the effluent was 13.2 ± 2.1 mg/L, with average influent concentrations of 27.9 ± 3.9 mg/L.

The dates where little ammonia was removed (Figure 5.11), combined with the extremely low nitrate (NO_3-N) effluent levels (Figure 5.12) on the same sampling dates suggests that nitrification in the MBR ceased from time to time

and that only organic nitrogen was being removed by the membrane separation process, presumably being incorporated in or forming solids. The small amount of NH_3-N that was removed could be the result of volatilization or sorption to organic particles.

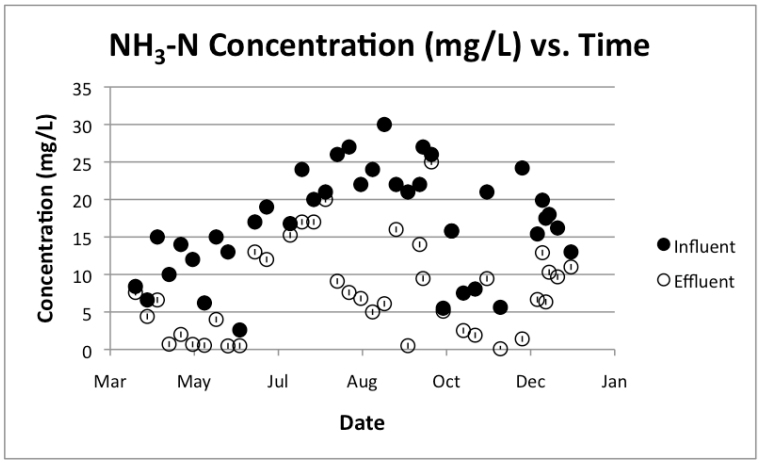


Figure 5.11. MBR Ammonia-Nitrogen

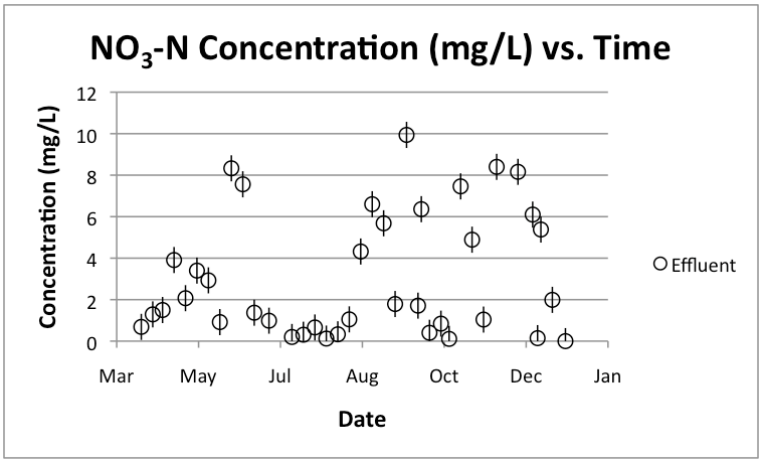


Figure 5.12. MBR Nitrate-Nitrogen

At times nitrogen removal seemed to severely decrease, particularly after recovery cleaning of the membrane. While the membrane was cleaned outside of the reactor, a process that often exceeded 24 hours in total duration, the mixed liquor received continued aeration without influent flow. Therefore, it is possible that decreased nitrification immediately after cleanings (July 9 and October 9) were the result of starvation and endogenous decay of autotrophic organisms. Another possibility is that the majority of viable autotrophic organisms were located on or in the caking on the membrane surface. Regardless, heterotrophic microbes are thought to recover much more quickly, while autotrophic nitrifiers are suspected to take as long as 7-10 days to recover. This seems to be reflected in the data. Furthermore, it was unexpected that some nitrification seemingly continued into the winter months as the literature reports nitrification as being minimal once reactor temperatures drop below 10°C (Tchobanoglous, Burton et al. 2003). The mean ammonia-nitrogen concentration in the effluent was 8.06 ± 2.03 mg/L, with an average influent concentration of 16.8 ± 2.3 mg/L.

Influent NO_3-N sampling and analysis was conducted in mid-June to allow more accurate nitrogen mass balances. Influent NO_3-N is commonly assumed to be zero. The influent concentration of NO_3-N remained consistently less than 1 mg/L upon several samplings, so influent testing was discontinued, the influent NO_3-N was assumed to be 0 mg/L, and this data is not illustrated in Figure 5.12. The trend in NO_3-N production is directly correlated to ammonia removal. High nitrate-nitrogen concentrations are seen during times when significant ammonia-

nitrogen removal occurred. The mean nitrate-nitrogen concentration in the effluent was 3.13 ± 0.99 mg/L.

While we suspect some phosphate removal may be do to biological uptake, the majority is probably due to phosphate in organic particles removed via solid/liquid separation of the membrane. A phosphate reagent shortage was experienced for three weeks in August, but once tests resumed it became apparent that removal rates had not changed much from before the cleaning on 9 July. Average effluent concentration was 6.34 ± 0.96 mg/L, as shown in Figure 5.13.

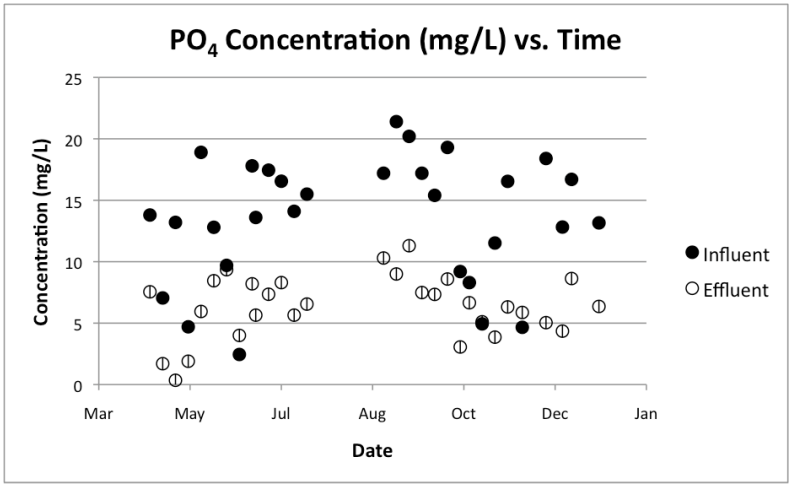


Figure 5.13. MBR Phosphate Concentrations

Table 5.1 on recaps average influent and effluent concentrations for the constituents tested during the entire MBR test period, with variance measured at

a 95% confidence level. Table 5.2 gives average influent, effluent and reactor conditions for the MBR between May 24 and September 22 of 2009 for the purposes of comparison to the bMBR, which was operated over the same timeframe in 2010. Please note MBR temperature was not recorded until June 24, 2009. Mass and volumetric removal will be discussed in Section 5.3.

Table 5.1. Average Influent, Effluent, and Reactor Conditions for the MBR (entire operational period)

	Average Influent Concentration (mg/L)	Average Effluent Concentration (mg/L)	Average Reactor Value (mg/L, °C)
TSS	170 ± 20	0.8 ± 0.2	
VSS	150 ± 20	0.7 ± 0.2	
Turbidity (NTU)	170 ± 20	0.9 ± 0.3	
COD	390 ± 40	15 ± 2	
BOD	110 ± 20	1.7 ± 0.6	
TN	28 ± 4	13 ± 2	
NH_3-N	17 ± 2	8.1 ± 2.0	
NO_3-N	0 (assumed)	3.1 ± 1.0	8.6 ± 2.1 (created)
PO_4-P	14 ± 2	6.3 ± 1.0	
pH	7.3 ± 0.2	7.6 ± 0.4	
D.O.			4.5 ± 0.8
MLSS			40 ± 12 (Surface) 1,700 ± 400 (Bottom) 8,800 ± 800 (Waste)

Table 5.1. Average Influent, Effluent, and Reactor Conditions for the MBR (entire operational period) Continued

	Average Influent Concentration (mg/L)	Average Effluent Concentration (mg/L)	Average Reactor Value (mg/L, °C)
MLVSS			33 ± 9 (Surface) 1,400 ± 300 (Bottom) 6,500 ± 500 (Waste)
Temperature (°C)			19 ± 3

Table 5.2. Average Influent, Effluent, and Reactor Conditions for the MBR (5/24/2009-9/22/2009)

	Average Influent Concentration (mg/L, NTU)	Average Effluent Concentration (mg/L, NTU)	Average Reactor Value (mg/L, °C)
TSS	180 ± 30	0.6 ± 0.2	
VSS	160 ± 30	0.5 ± 0.1	
Turbidity (NTU)	180 ± 30	0.6 ± 0.1	
COD	430 ± 60	16 ± 3	
BOD	110 ± 20	1.5 ± 0.7	
TN	30 ± 6	15 ± 3	
NH_3-N	19 ± 4	9.4 ± 3.5	
NO_3-N	0 (assumed)	2.7 ± 1.5	10 ± 4 (created)
PO_4-P	15 ± 3	7.7 ± 1.3	
pH	7.2 ± 0.5	7.4 ± 0.8	
D.O.			2.9 ± 0.5

Table 5.2. Average Influent, Effluent, and Reactor Conditions for the MBR (5/24/2009-9/22/2009) Continued

	Average Influent Concentration (mg/L)	Average Effluent Concentration (mg/L)	Average Reactor Value (mg/L, °C)
MLSS			46 ± 21 (Surface) 3,000 ± 700 (Bottom) 9,800 ± 1,200 (Waste)
MLVSS			37 ± 15 (Surface) 2,400 ± 500 (Bottom) 6,900 ± 800 (Waste)
Temperature (°C)			26 ± 2

5.2 bMBR DATA

Operation of the new reactor began May 21st, 2010. Suspended solids concentrations (Figure 5.14) in the influent were inconsistent due to diurnal variations and inflow and infiltration into the collection system. Error bars were calculated to be ± 50 mg/L for the influent and ± 1 mg/L for the effluent. However, the constituent concentrations in the influent during the summer of 2010 seemed to be more dilute than during the same period in 2009. Effluent concentrations of suspended solids remained consistently low throughout the testing period, suggesting the extreme operating conditions in the winter of 2009-2010 did not cause critical harm to the membrane. Mean TSS concentrations

over the operational period were 121 ± 17 mg/L for the influent and 0.60 ± 0.25 mg/L in the effluent.

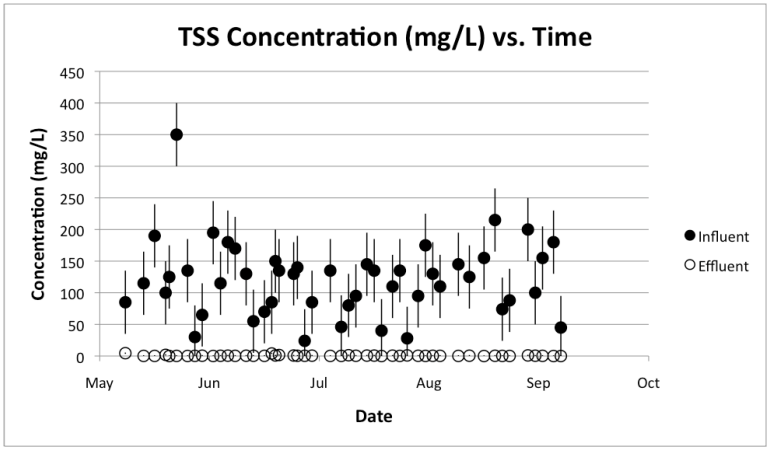


Figure 5.14. bMBR Total Suspended Solids

Variations in volatile suspended solids concentration (Figure 5.15) of the reactor's effluent and influent seemed well associated to total suspended solids concentrations, and often constituted the vast percentage of total suspended solids. Error bars were determined to be the same as those for TSS. The supply of glass filters used for measuring TSS and VSS was exhausted on September 12th, and the new brand has a smoke point well below the 550°C needed for a standard VSS test. Therefore, results after September 12th have been omitted for VSS and MLVSS (Figures 5.15 and 5.18, respectively). Mean VSS concentrations over the operational period were 109 ± 16 mg/L for the influent and 0.45 ± 0.18 mg/L in the effluent.

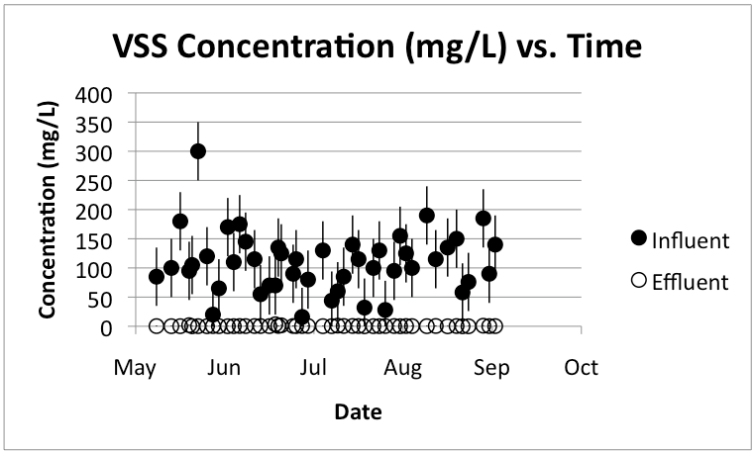


Figure 5.15. bMBR Volatile Suspended Solids

The bMBR began operation with approximately 20 gallons of seed sludge from the Deployable Baffled Bioreactor (dBBR), which was being tested in close proximity to this reactor, and 10 gallons of plant influent. The anoxic/aerobic weir was set to allow for approximately 1 L/min flow over the top. Such a setting, when the effluent pump was set to draw 0.33 mL/min in permeate, should induce a return rate to the anoxic zone of 0.66 mL/min. The aerobic/anoxic baffle was set to allow for ¼" clearance between the rubber flap on the baffle and the bottom of the reactor.

Sludge wasting was performed on Mondays, Wednesdays and Fridays with volumes of 9 L, 6 L, and 6 L, respectively. This resulted in an estimated sludge retention time (SRT) of 41 days. Actual SRT was probably less due to physical removal of floating and foaming sludge as it occurred, as well as periodic scouring of the membrane to decrease fouling and TMP. Initially, MLSS in the reactor appeared much more evenly distributed than in the MBR, with

concentrations in both the aerobic and anoxic zones exceeding 1 g/L after the first week (Figure 5.16). This higher concentration of biomass seemed to foul the reactor rather quickly, as TMP would generally increase to maximum permeate pressure within a week.

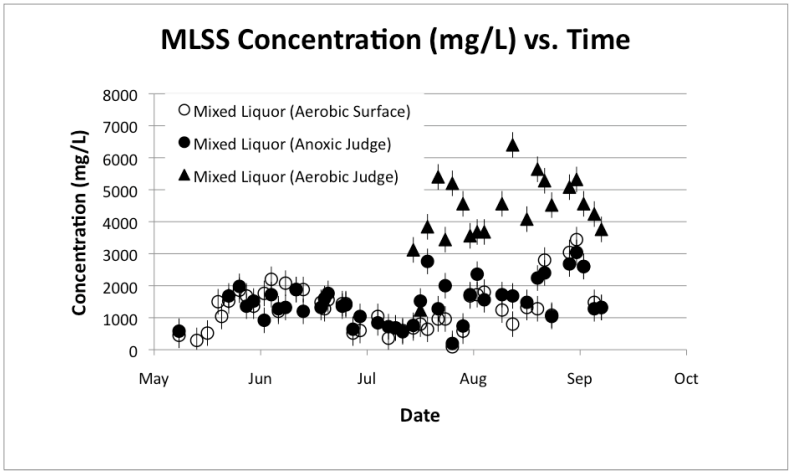


Figure 5.16. bMBR Mixed Liquor Suspended Solids

To maintain desired TMP, the membrane was removed from the reactor and scoured with tap water nearly every other day. This was necessary to continue membrane operation, but occurred at the expense of reactor biomass that was sloughed off in the process. Much of this biomass was concentrated in and around the individual hollow-fiber membranes. The outer portion appeared to be tannish-brown, while the inner layers were dark brown to black in appearance. The black biomass would often fall off in flakes, perhaps suggesting significant compression against the membrane surface without the

presence of oxygen had occurred. Also, this reactor seemed to have significant large-bubble foaming in the aerobic zone, which would accumulate as fine-bubble foam or scum in the anoxic zone. These layers were occasionally removed to allow for proper sampling and testing of the reactor. However, such maintenance may have caused an overall decrease in SRT, but probably not as significantly as the effect of membrane scouring on SRT.

Upon start up, the dissolved oxygen in the reactor was very high in both the aerobic and anoxic zones (Figure 5.17).

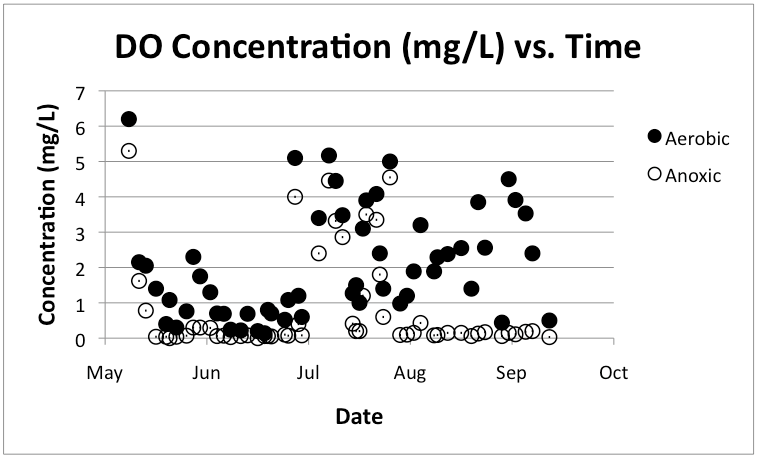


Figure 5.17. bMBR Dissolved Oxygen

Gradually, both began to decrease in concentration while the disparity between the zones increased. However, by June 7th D.O. in the aerobic zone had fallen below 1 mg/L, which was believed to be significantly affecting nitrification. In response, the aerobic/anoxic baffle was lowered so that the

rubber flap on the baffle was barely touching the bottom of the reactor. It was thought that increased HRT in the aerobic zone would increase D.O. through contact time. Mean D.O. concentrations over the operational period were 2.04 ± 0.42 mg/L in the aerobic zone and 0.82 ± 0.38 mg/L in the anoxic zone.

This assumption appeared to be correct at first as D.O. climbed back above 2 mg/L, but within a week it had fallen back to near 1 mg/L. Shortly thereafter, the sludge judge revealed that there was sludge accumulating in the "dead" zones within the aerobic portion of the reactor. Therefore, increasing aerobic/anoxic baffle depth may have led to decreased mixing in the aerobic zone as the liquor was not ushered through the zone as quickly as before. Biomass build-up in portions of the aerobic zone may be demonstrated by a decrease in concentration of the aerobic surface and composite anoxic samples as seen in Figures 5.16 and 5.18.

On July 29th the baffle was raised to its original position, the number of rain events began to decrease, and biomass concentrations throughout the aerobic zone were taken and analyzed using the sludge judge for a composite sample. The difference in concentrations can be seen in Figures 5.16 and 5.18. MLVSS (Figure 5.18) tracked well with MLSS (Figure 5.16), with non-volatile suspended solids comprising less than 20% of MLSS concentration. Error bars for MLSS and MLVSS were calculated to be ± 400 mg/L for all three zones. Mean concentrations of the MLSS over the operational period of the bMBR in the aerobic zone via surface grab sample, anoxic zone via sludge judge, and in the aerobic zone using a sludge judge were 1,330 ± 210 mg/L, 1,480 ± 200 mg/L,

and 4,340 ± 500 mg/L, respectively. Mean concentrations of MLVSS concentrations in the same zones were 1,160 ± 192 mg/L, 1,310 ± 177 mg/L, and 3,540 ± 430 mg/L, respectively.

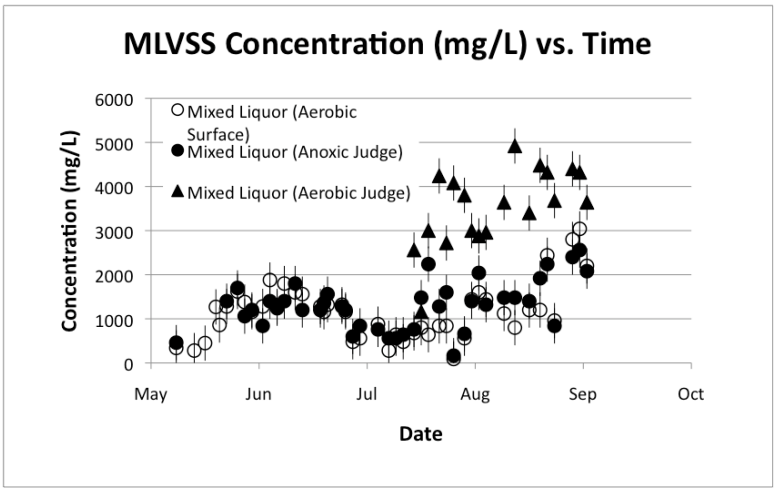


Figure 5.18. bMBR Mixed Liquor Volatile Suspended Solids

Influent COD concentration (Figure 5.19) was relatively dilute through the summer of 2010 compared to the data from 2009 shown in Figure 5.8. The dilute nature of the 2010 influent seems to provide evidence of reactor efficacy. High concentrations of biomass were retained in the reactor, significant denitrification was accomplished, and overall COD removal was high. Mean COD concentrations over the operational period were 308 ± 41 mg/L in the influent and 14.4 ± 2.0 mg/L in the effluent.

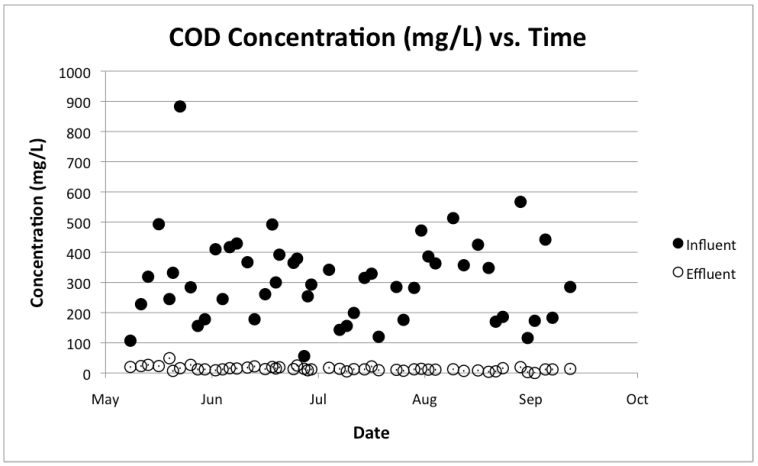


Figure 5.19. bMBR Chemical Oxygen Demand

Interestingly, total nitrogen concentrations during the summer of 2010 (Figure 5.20) seemed to be almost as concentrated as those shown in 2009 (Figure 5.10).

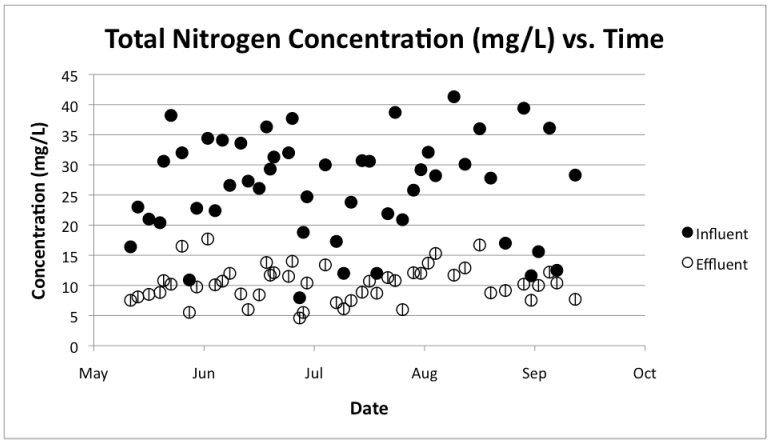


Figure 5.20. bMBR Total Nitrogen

Since we would normally expect to see equivalent dilution of COD and TN, it may be assumed that there was either a decrease in COD input or an increase in TN input somewhere in the collection system. However, it may also be said that the number of measurements made in 2009 was only one-third of those in 2010. Mean TN concentrations over the operational period were 26.3 ± 2.5 mg/L in the influent and 10.3 ± 0.9 mg/L in the effluent.

As noted in Figure 5.17, D.O. concentration was near 1 mg/L before the installation of the larger blower, and almost always near or greater than 2 mg/L after installation of the larger blower on July 15. It was suspected that complete nitrification was being inhibited by low D.O. concentrations. This hypothesis seems to have been correct as displayed by ammonia-nitrogen concentrations shown in Figure 5.21.

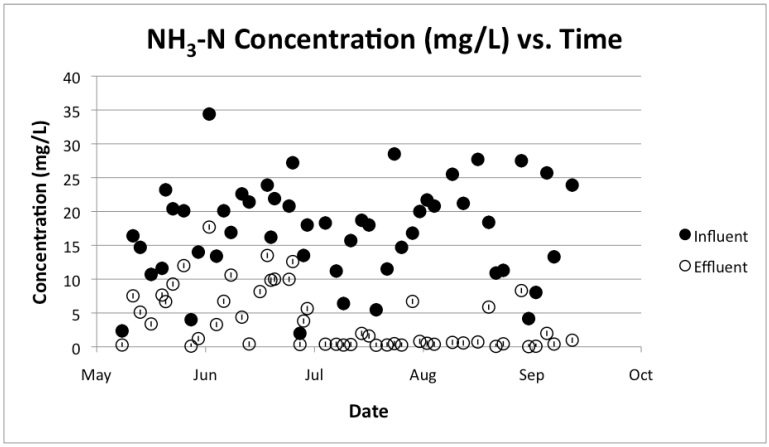


Figure 5.21. bMBR Ammonia-Nitrogen

Aside from three outliers, effluent ammonia-nitrogen concentrations were always less than 2 mg/L after installation of the fine-bubble diffuser and the first recovery cleaning. In fact, after July 16th (when operation recommenced after cleaning and large blower installation) average ammonia-nitrogen removal was 92%. Mean NH_3-N concentrations over the operational period were 17.1 ± 2.1 mg/L in the influent and 4.02 ± 1.27 mg/L in the effluent.

Almost all of the aforementioned outliers in effluent ammonia-nitrogen can be attributed to D.O. deficiency in the aerobic zone. To encourage maximum nitrification and denitrification in the separate zones, D.O. in the aerobic zone was targeted to be as close to 2 mg/L as possible. Conversely, D.O. in the anoxic zone was targeted to be as close to 0 mg/L as possible. To achieve these objectives, air flowrate to the fine-bubble diffusers was controlled via bleed-off valve. Although somewhat inaccurate in and of itself, this method proved somewhat successful in managing D.O. concentration differential between the zones. Unfortunately, temperature and influent concentration differences between hours and days negatively affected the pursuit of peak reactor performance. That is, ideal airflow at a specific time on a specific day was unique to that point in time. Constant D.O. monitoring with PLC control would allow for ideal conditions to consistently be met.

Effluent concentrations of nitrate-nitrogen (Figure 5.22) were generally below the 10 mg/L limit for losing streams and drinking water as required by the US EPA. The D.O. in the anoxic zone was less than 1 mg/L (Figure 5.17) for significant periods of time before and after July 15. However, before large blower

installation, the D.O. in the aerobic zone was near 1 mg/L on a consistent basis. Such a low D.O. may have encouraged some denitrificaiton within the flocs or areas of concentrated biomass within the aerobic zone. Also, after D.O. in the aerobic zone was consistently maintained at or above 2 mg/L, the nitrification rate increased in the reactor. Increased nitrate-nitrogen concentrations in the effluent after July 16 may be the result of great loading along with, or instead of, a decrease in denitrification rate. Mean NO_3-N concentrations in the effluent over the operational period were 4.32 ± 1.04 mg/L.

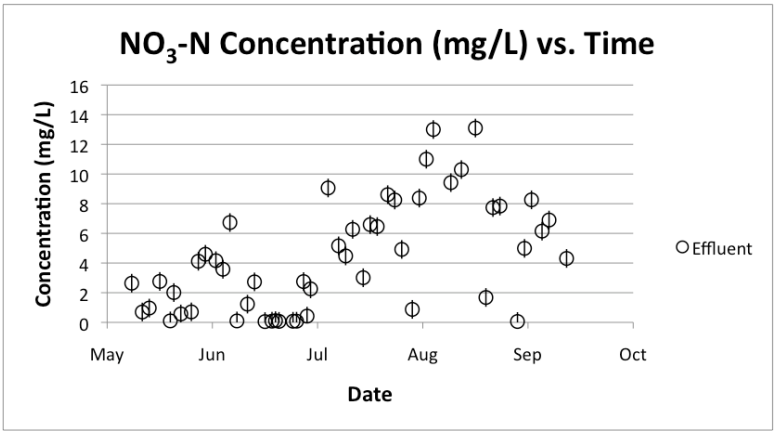


Figure 5.22. bMBR Nitrate-Nitrogen

Looking at Figure 5.22, it appears that the instances where effluent nitrate-nitrogen exceeded 10 mg/L were during times when nitrate-nitrogen production (via nitrification) was greatest. Nitrite-nitrogen was not analyzed during this experiment and was assumed to be negligible. Furthermore, it was assumed

that all nitrate-nitrogen in the reactor was produced via nitrification of ammonia-nitrogen.

Temperature of the bMBR is illustrated in Figure 5.23. Mean reactor temperature over the operational period was 28.2 ± 0.7°C. While temperature appears to have increased by 5-6 degrees over the first weekend in August, a seemingly unexplainable occurrence happened as well.

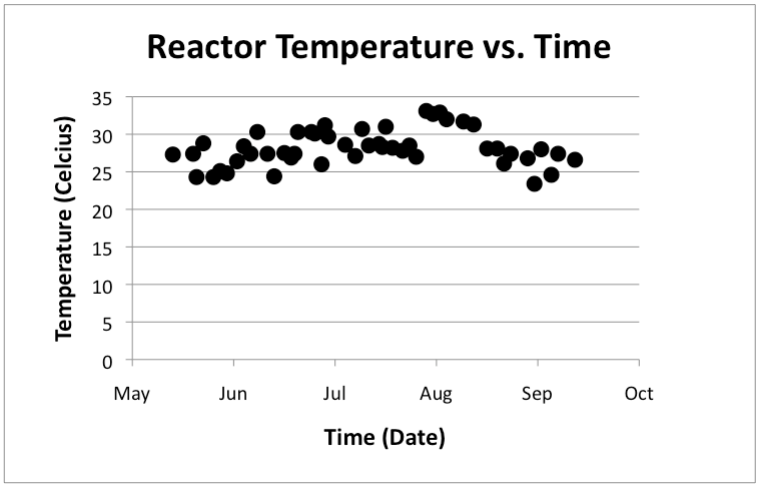


Figure 5.23. bMBR Temperature

On August 6th the MLSS in the reactor appeared to be extremely dilute upon arriving on site at 10 a.m. Analysis suggested the surface aerobic MLSS had dropped 90% from approximately 900 mg/L to 90 mg/L (Figure 5.16); at this time the D.O. was also noted to be high at 5.0 mg/L in the aerobic zone and 4.6 mg/L in the anoxic zone. TMP on the membrane was near the maximum of

seven inches of Hg. Upon returning to the site at 6 pm, the MLSS still appeared very dilute, but TMP had increased to approximately 12" Hg and was dangerously close to exceeding safe operational parameters of the membrane. The membrane was removed from the reactor and scoured using tap water. A significant amount of sludge was sloughed from the membrane module (estimated to be 1-2 lbs), and the module was returned to the reactor.

Once operation commenced the MLSS appeared anaerobic – that is, it was very black in color and had an unpleasant aroma. Transmembrane pressure had decreased back to approximately 7" Hg. By August 9th operation appeared to be normal with aerobic D.O. near 1 mg/L and anoxic D.O. near 0 mg/L. This incident did not repeat itself and proper environmental conditions were maintained in the aerobic and anoxic zones for the rest of the test period. Low aerobic D.O. after this incident was occasionally seen, but always as the result of clogged coarse-bubble diffusers or too much bleed-off from the fine-bubble diffusers.

The stoichiometry of nitrification (Equation 3) results in two protons generated for every one molecule of ammonium that undergoes nitrification. Likewise, during denitrification (Equation 4) one molecule of hydroxide is generated for every molecule of nitrate that undergoes denitrification. However, for every ten molecules of nitrate that undergo denitrificaiton, one molecule of ammonia is formed. With reactor pH generally above 7 (Figure 5.24), the ammonia would quickly ionize to ammonium – thus removing one proton per ammonia. Therefore, the relatively steady pH in the bMBR would seem to

suggest that a significant amount of denitrification is taking place. The municipal drinking water in this area has significant alkalinity, which may contribute to pH buffering within the reactor. It is possible that without such high alkalinity the effluent pH would be significantly lower. The membrane could act to sieve a portion of $CaCO_3$, which is retained in the reactor and counteracts some of the effects of nitrification. Mean pH over the operational period was 7.17 ± 0.05 in the influent and 7.38 ± 0.04 in the effluent.

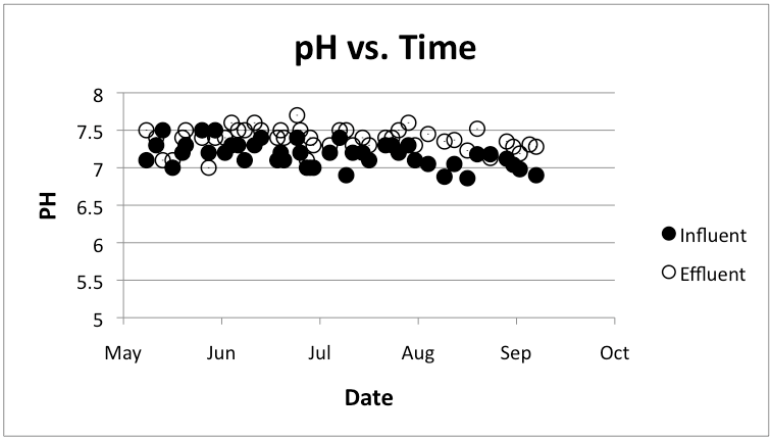


Figure 5.24. bMBR pH

Table 5.3 shows means of influent, effluent and reactor data, with variance calculated at a 95% confidence level. Volumetric and mass removal data will be addressed in Section 5.3. Reactor values for MLSS and MLVSS are recorded as aerobic zone from surface, anoxic zone with sludge judge, and aerobic zone using sludge judge (SJ).

Table 5.3. Average Influent, Effluent, and Reactor Conditions for the bMBR

	Average Influent Concentration (mg/L, NTU)	Average Effluent Concentration (mg/L, NTU)	Average Reactor Value (mg/L, °C)
TSS	120 ± 20	0.6 ± 0.3	
VSS	110 ± 20	0.5 ± 0.2	
Turbidity (NTU)	N/A	N/A	
COD	310 ± 40	14 ± 2	
BOD	N/A	N/A	
TN	26 ± 3	10 ± 1	
NH_3-N	17 ± 2	4.0 ± 1.3	
NO_3-N	0 (assumed)	4.3 ± 1.0	13 ± 2
PO_4-P	N/A	N/A	
pH	7.2 ± 0.1	7.4 ± 0.0	
D.O.			2.0 ± 0.4 (Aerobic) 0.82 ± 0.38 (Anoxic)
MLSS			1,300 ± 200 (Aerobic) 1,500 ± 200 (Anoxic) 4,300 ± 500 (SJ)
MLVSS			1,200 ± 200 (Aerobic) 1,300 ± 200 (Anoxic) 3,500 ± 400 (SJ)
Temperature (°C)			28 ± 1

Table 5.4. Average Influent, Effluent, and Reactor Conditions for the bMBR After Large Blower Installation on July 15

	Average Influent Concentration (mg/L, NTU)	Average Effluent Concentration (mg/L, NTU)	Average Reactor Value (mg/L, °C)
TSS	120 ± 20	0.39 ± 0.14	
VSS	110 ± 19	0.33 ± 0.12	
Turbidity (NTU)	N/A	N/A	
COD	290 ± 50	11 ± 2	
BOD	N/A	N/A	
TN	26 ± 4	10 ± 1	
NH_3-N	17 ± 3	1.3 ± 0.9	
NO_3-N	0 (assumed)	6.8 ± 1.3	15 ± 3
PO_4-P	N/A	N/A	
pH	7.1 ± 0.1	7.4 ± 0.1	
D.O.			2.7 ± 0.5 (Aerobic) 1.0 ± 0.5 (Anoxic)
MLSS			1,300 ± 400 (Aerobic) 1,600 ± 300 (Anoxic) 4,300 ± 500 (SJ)
MLVSS			1,200 ± 340 (Aerobic) 1,400 ± 300 (Anoxic) 3,500 ± 400 (SJ)
Temperature (°C)			29 ± 1

5.3 DATA COMPARISON

The MBR was operated in 2009 and the bMBR in 2010. A direct data comparison is not necessarily indicative of relative reactor efficacy due to differences in temperature, influent constituent concentration and other factors. Simultaneous testing was not possible due to limited availability of membrane modules, but for the purposes of comparison only data from the summer of 2009 will be considered for the MBR. The ideal situation would have been to test the bMBR as designed, an MBR whose volume was equal to the aerobic volume in the bMBR, and a MBR whose volume was equal to the total reactor volume of the bMBR.

Even though the ideal criteria were not met for these experiments, comparison of the two reactor configurations can still be made. First, knowing the flow rate through the reactors as well as the influent and effluent concentrations for each constituent, we can determine the volumetric removal rate expressed in mg/L*day. Next, we can look at the mass per time loading rate for each unit of volume in the reactor and juxtapose it with the removal mass per time for each unit of volume in the reactor. Complete removal (e.g., 100%) can be plotted as a 1:1 relationship and relative observations between reactors can be made.

Figures 5.25 and 5.26 show volumetric removal of COD in the bMBR and MBR, respectively. Volumetric removal in the bMBR seems to be much more consistent, while removal in the MBR seems to be greater. However, influent

COD concentrations were significantly higher during 2009. Not knowing this COD discrepancy might lead one to misinterpret this comparison by concluding that removal rates were higher due to the configuration of the reactor.

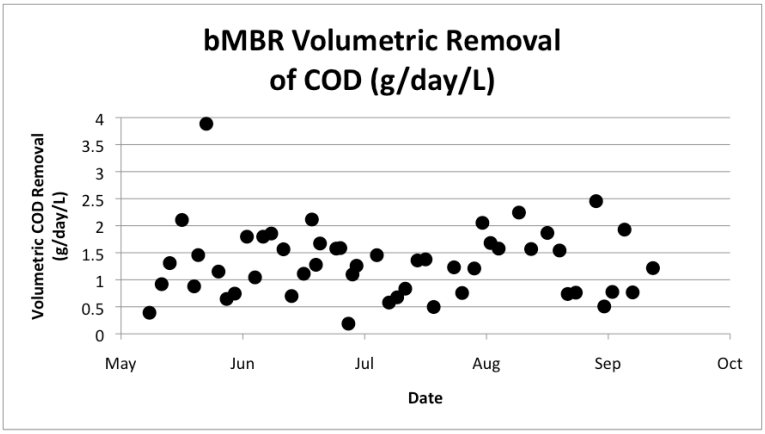


Figure 5.25. Volumetric Removal of COD in bMBR

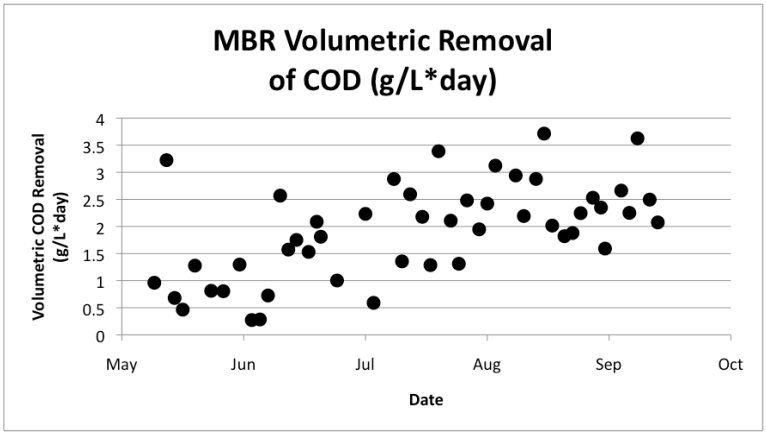


Figure 5.26. Volumetric Removal of COD in MBR

Solid/liquid separation by membrane filtration probably contributed to the high removal rate vs. loading rate as seen in Figures 5.27 and 5.28 for the bMBR and MBR, respectively.

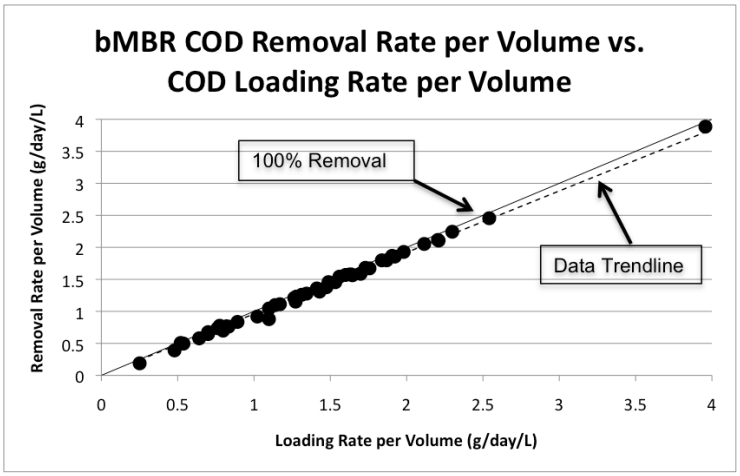


Figure 5.27. Volumetric Removal Rate vs. Loading Rate of COD in bMBR

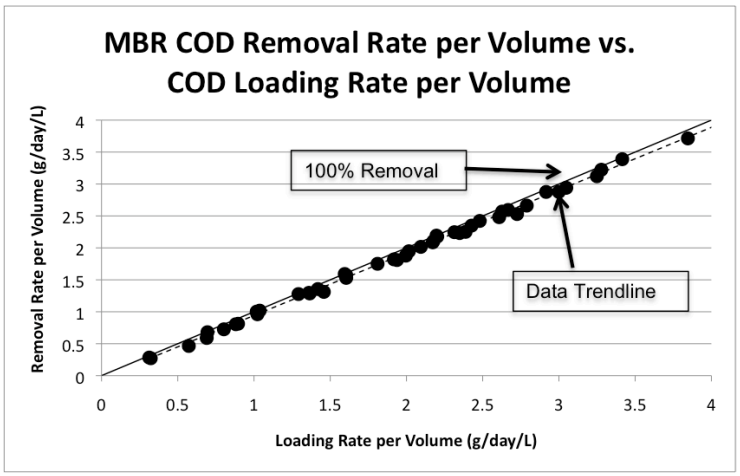


Figure 5.28. Volumetric Removal Rate vs. Loading Rate of COD in MBR

Both reactors show consistently high removal rates suggesting that the different configurations had little to no effect on overall reactor performance in removing COD. The line representing 100% removal, is plotted as a solid line to give a point of reference. A dashed line is plotted that shows the data trend.

Volumetric removal of total nitrogen in the bMBR and MBR can be viewed in Figures 5.29 and 5.30, respectively.

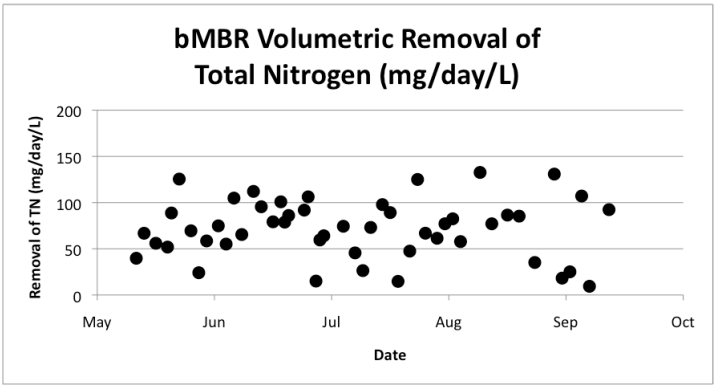


Figure 5.29. Volumetric Removal of Total Nitrogen in bMBR

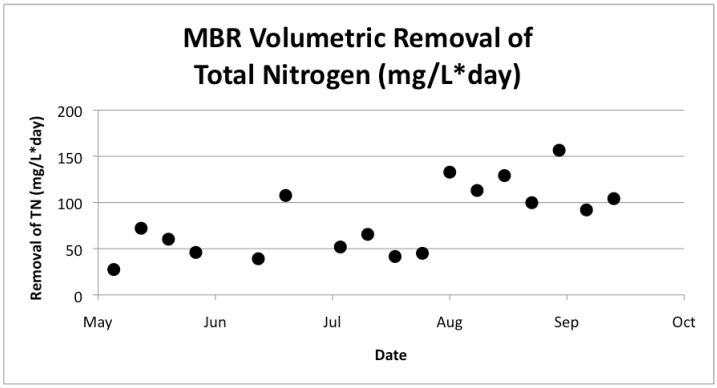


Figure 5.30. Volumetric Removal of Total Nitrogen in MBR

Although there are fewer data points during the MBR test period (Figure 5.30) it seems that volumetric removal was less than in the bMBR (Figure 5.29).

Volumetric loading and removal was calculated using total reactor volumes of 98 L for the MBR and 107 L for the bMBR. The larger overall volume of the bMBR decreases the volumetric removal per liter. Also, as with COD, total nitrogen concentrations were higher in 2009 than in 2010. Subtracting the mean ammonia-nitrogen concentration from the mean total nitrogen concentration, and assuming nitrate and nitrite concentrations are negligible, leads to the determination that the MBR had organic nitrogen concentrations that were roughly 20% higher than that of the bMBR. Therefore, the mere physical separation of organic particles from the liquid via membrane filtration could be interpreted as increased total nitrogen removal in the MBR. It may be that there was greater biological degradation in the bMBR relative to the MBR than what this analysis suggests.

Figure 5.31 displays the removal rate per volume versus the loading rate per volume for the bMBR. The data points seem to trend nicely with the 100% removal line even though they fall below it. Compare this with Figure 5.32 for the MBR. The data points in Figure 5.32 appear to be more random and appear to diverge more from the 100% removal line than those in Figure 5.31. Thus, the conclusion can be made that the bMBR not only had a better removal rate for total nitrogen than did the MBR, but was also more consistent in the rate with which nitrogen was removed throughout the testing period. Finally, the loading

rate of the bMBR (Figure 5.31) appears to trend lower than that of the MBR (Figure 5.32).

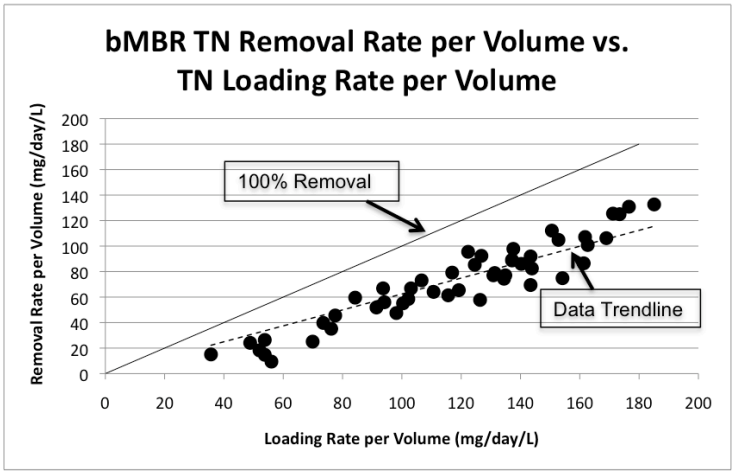


Figure 5.31. Volumetric Removal Rate vs. Loading Rate of Total Nitrogen in bMBR

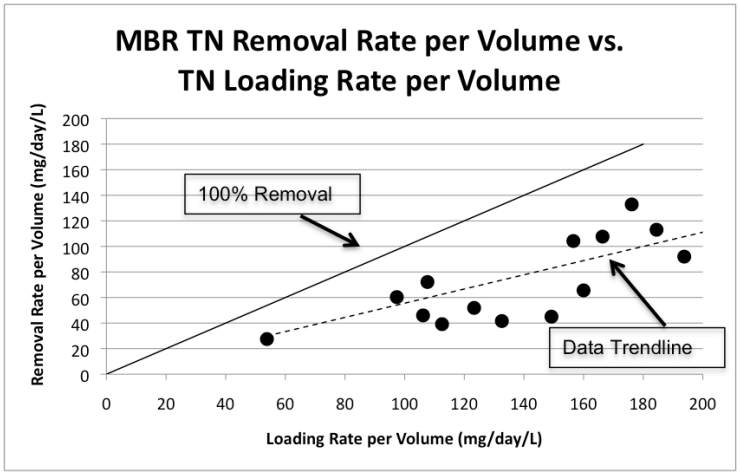


Figure 5.32. Volumetric Removal Rate vs. Loading Rate of Total Nitrogen in MBR

This supports the theory that the apparent similarities in removal rates and percentage removal within the reactors may be misleading due to the possibility of increased organic nitrogen concentrations on the MBR being removed through physical separation rather than biological degradation, although both were to have likely occurred. It could be that different mechanisms were at work in the reactors, which contributed to similar results.

Ammonia-nitrogen being a part of the calculation for total nitrogen, it should be of little surprise that the ammonia-nitrogen concentrations also seemed to be higher in 2010 than in 2009. The volumetric removal of ammonia-nitrogen in the bMBR (Figure 5.33) was greater than the removal rate in the MBR (Figure 5.34) due to the relative volumes of the reactors. Figure 5.35 shows volumetric removal for the bMBR if only the aerobic zone with a volume of 33 L is considered.

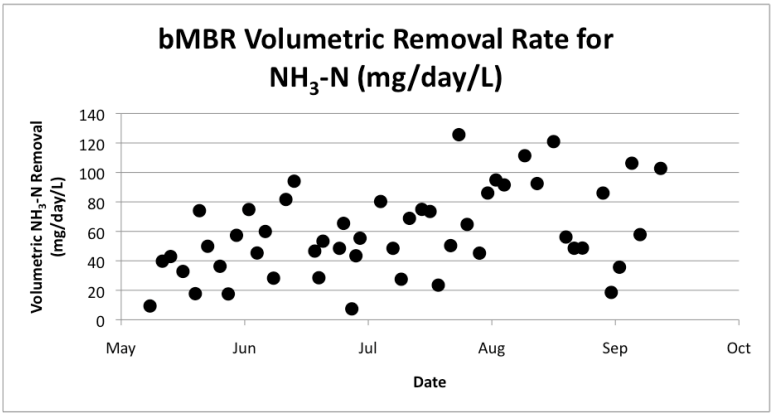


Figure 5.33. Volumetric Removal of Ammonia-Nitrogen in bMBR

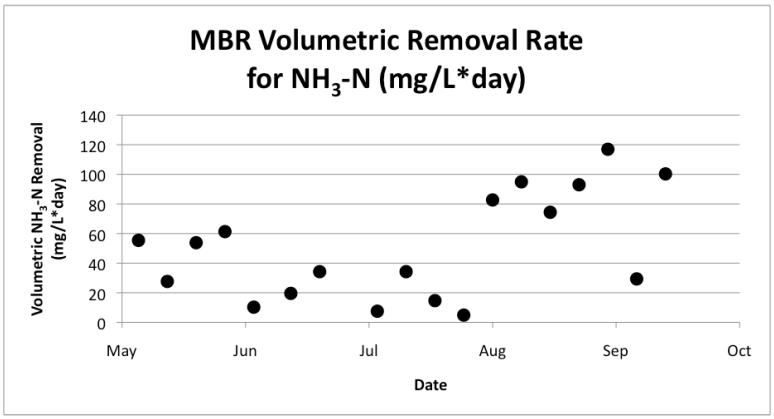


Figure 5.34. Volumetric Removal of Ammonia-Nitrogen in MBR

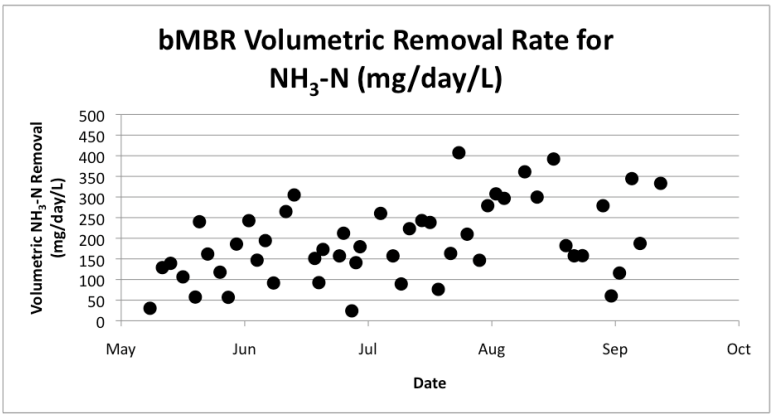


Figure 5.35. Volumetric Removal of Ammonia-Nitrogen in the Aerobic Zone of the bMBR

The aerobic zone in the bMBR with a volume of 33 L may be a better consideration for this calculation since it is well established that the vast majority of nitrification takes place under aerobic conditions. Conversely, it is difficult to determine what the active aerobic volume was in the MBR. While the entire reactor appeared to be aerobic based on D.O. concentrations, intuition suggests

otherwise due to some denitrificaiton. Therefore, the entire reactor volume of 98 L was used in this calculation.

Figures 5.36 and 5.37 demonstrate the removal rate of ammonia-nitrogen per volume versus the loading rate per volume for the bMBR and MBR, respectively. While both reactors appear to have achieved 100% (or very near 100%) removal on occasion, the bMBR seems to have done so much more frequently. Figure 5.38 shows the removal versus loading in the bMBR if only the aerobic volume is considered for the calculation. Since the difference in total volume to aerobic volume is 107 L to 34 L, or approximately 3:1, the rates in Figure 5.38 are approximately three-times those in Figure 5.36. However, there is no difference with respect to percent removal.

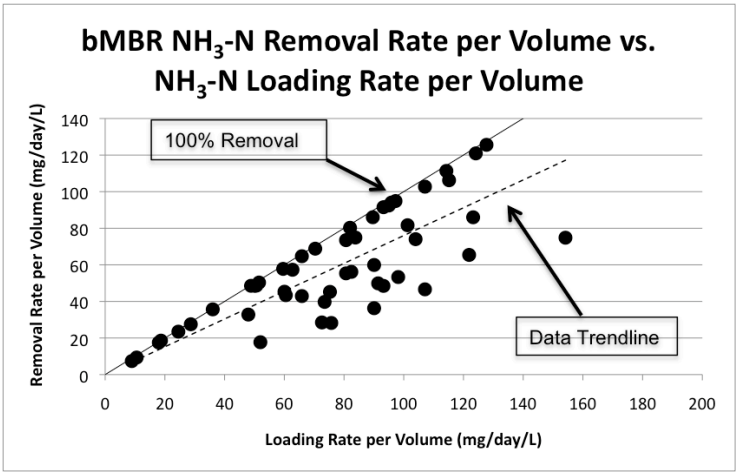


Figure 5.36. Volumetric Removal Rate vs. Loading Rate of Ammonia-Nitrogen in bMBR

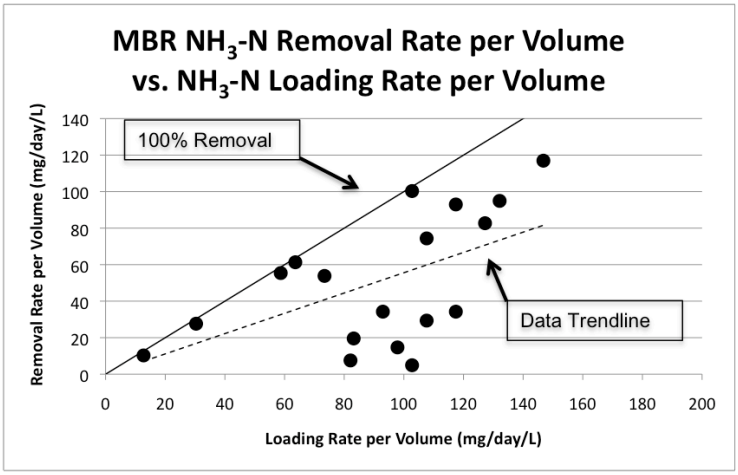


Figure 5.37. Volumetric Removal Rate vs. Loading Rate of Ammonia-Nitrogen in MBR

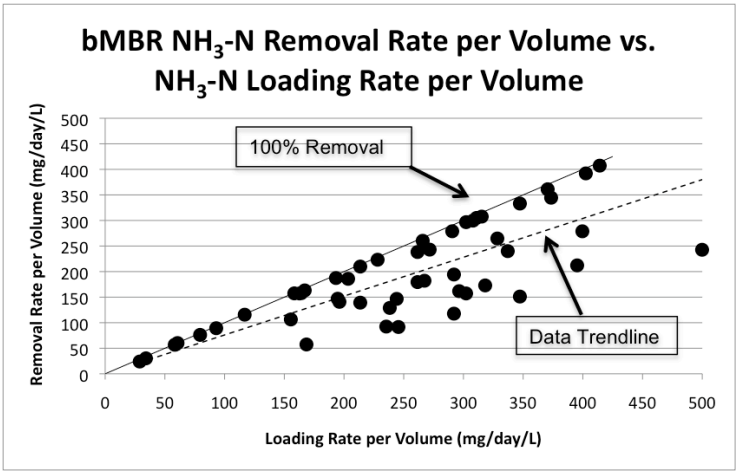


Figure 5.38. Volumetric Removal Rate vs. Loading Rate of Ammonia-Nitrogen in the Aerobic Zone of the bMBR

If we were to look at the same aerobic bMBR graph only after the large blower installation we get the data seen in Figure 5.39. This suggests that 90%+ nitrification rates are possible if the D.O. is maintained at a high enough concentration.

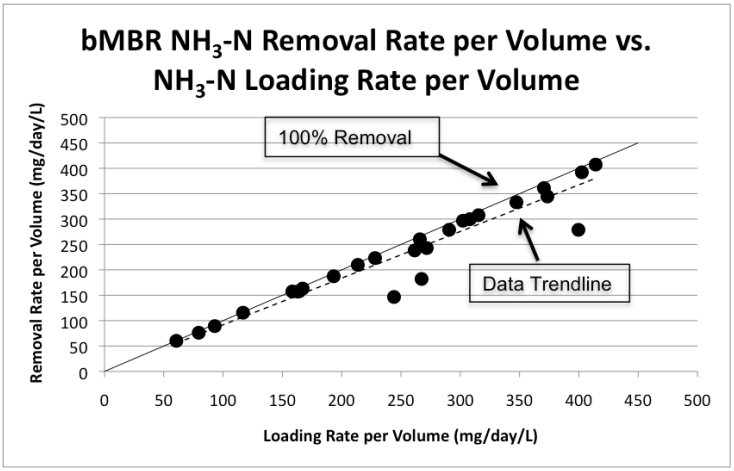


Figure 5.39 Volumetric Removal Rate vs. Loading Rate of Ammonia-Nitrogen in Aerobic Zone of the bMBR After Large Blower Installation

The three outliers in Figure 5.39 can be attributed to specific operational abnormalities such as clogged coarse-bubble diffusers and/or loose fine-bubble diffuser lines. Therefore, the bMBR appears to have been remarkably consistent and effective in removing ammonia-nitrogen when aerobic D.O. concentration at or above 2 mg/L is maintained.

Volumetric removal rates for nitrate-nitrogen in the bMBR appear to have been the same between the two reactors if total volumes are considered. Figure 5.40 shows bMBR removal rates of nitrate-nitrogen, while Figure 5.41 shows MBR removal rates. It appears that the values between bMBR and MBR volumetric removal of nitrate-nitrogen are roughly the same. Even though values in the MBR for the first part of the summer appear to be less, this may be because of fewer overall data points. The greater number of data points in the bMBR should result in a smaller variance.

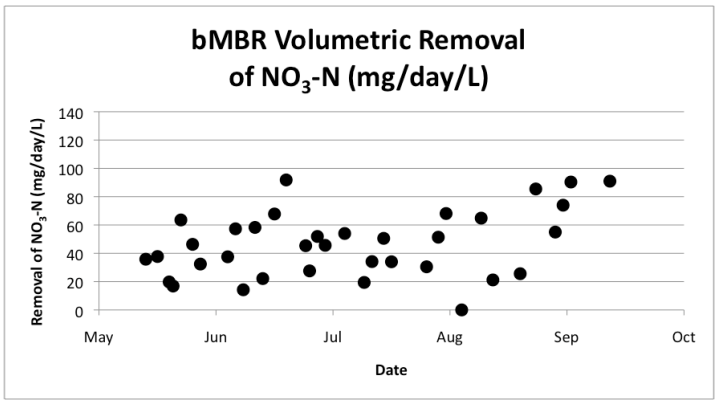


Figure 5.40. Volumetric Removal of Nitrate-Nitrogen in bMBR

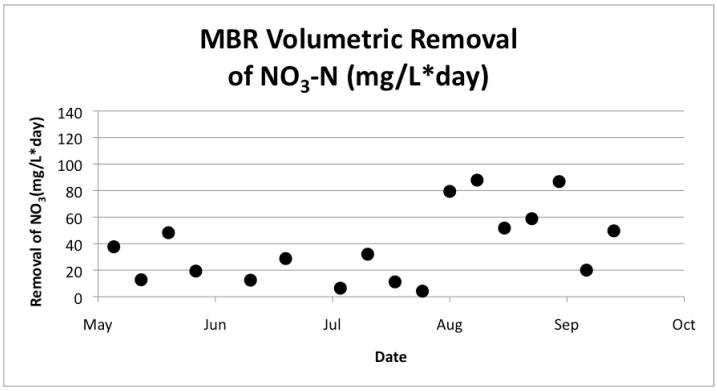


Figure 5.41. Volumetric Removal of Nitrate-Nitrogen in MBR

Figure 5.42 shows nitrate-nitrogen removal rates in the bMBR if only the anoxic volume of 74 L is considered. Since the total volume of the bMBR is 107 L, analyzing the data by only considering the anoxic zone in the calculation results in rates that are roughly 33% higher.

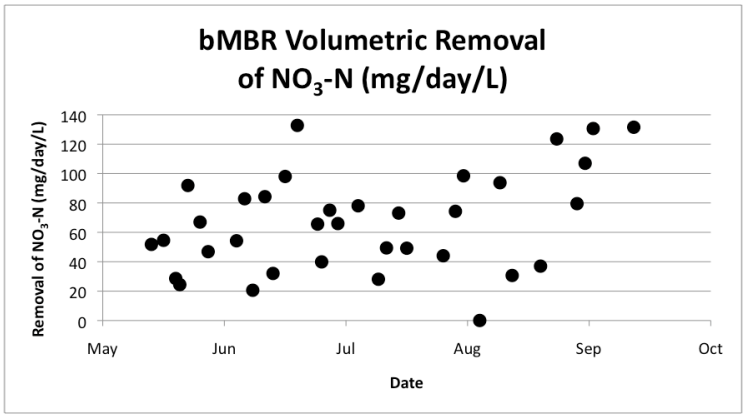


Figure 5.42. Volumetric Removal of Nitrate-Nitrogen in the Anoxic Zone of the bMBR

Since denitrification primarily takes place under anoxic conditions, it makes sense to use data from Figure 5.42 for kinetic calculations. The data in the figure suggests a greater volumetric removal rate in the anoxic zone of the bMBR than in the MBR. The bMBR had an anoxic volume of approximately 74 L, while the anoxic volume of the MBR is unknown. Therefore, it was assumed, perhaps poorly, that the entire reactor volume of 98 L was contributing to the denitrification process. While this assumption may or may not be poor, it is undeniable that unknown aerobic and anoxic volumes (i.e., simultaneous nitrification/denitrification) make modeling a reactor much more difficult than if these zones are well defined.

Figures 5.43 and 5.44 show nitrate-nitrogen removal rate per volume for the bMBR and MBR, respectively. While it appears that the bMBR achieves 100% nitrate-nitrogen removal more frequently than the MBR, the relative number of data points could be misleading. The bMBR had higher mass

loadings of nitrate-nitrogen due to greater capacity for nitrification than the MBR, which leads to higher potential removal rates. At the same time, percentage removal can stay the same or even decrease. While it's unknown how the removal rate in the MBR would respond to loading rates greater than 90 mg/day/L, it's unlikely that performance would be better than the bMBR for two primary reasons. First, defined zones in the bMBR allow for better control of D.O. concentration in the anoxic zone and thus limits its inhibitive affects. Second, the defined pre-anoxic zone allows for high BOD in the anoxic zone to maximize denitrification. Figure 5.45 shows this data for the bMBR if only the anoxic zone is considered.

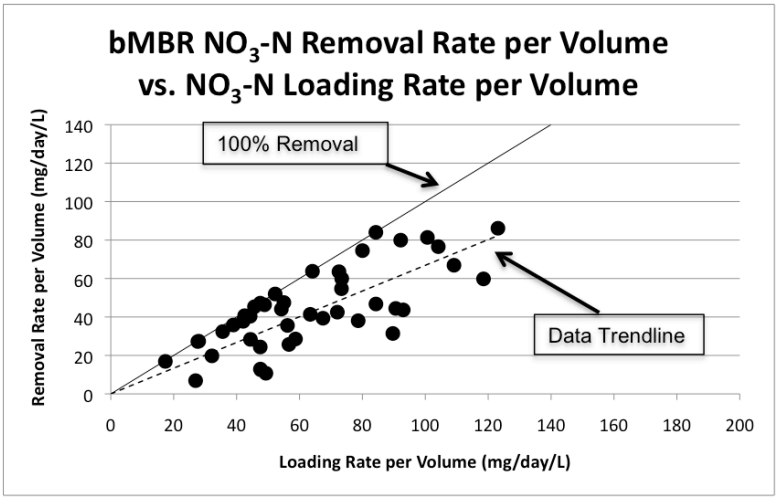


Figure 5.43. Volumetric Removal Rate vs. Loading Rate of Nitrate-Nitrogen in bMBR

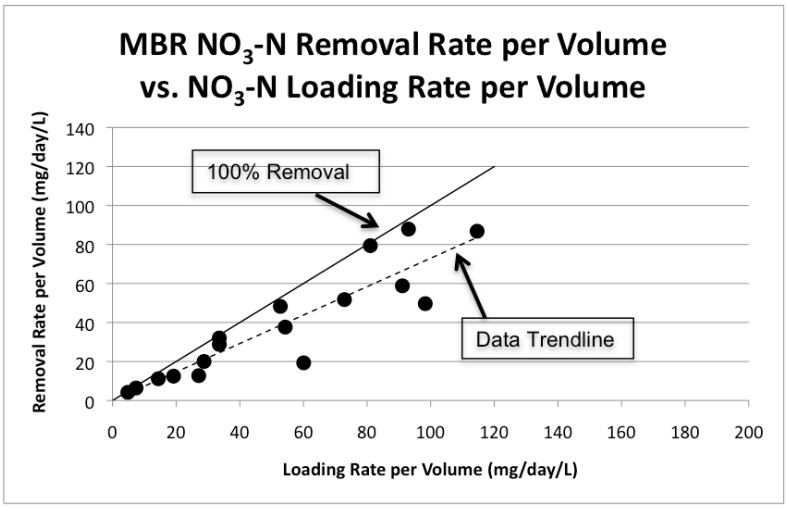


Figure 5.44. Volumetric Removal Rate vs. Loading Rate of Nitrate-Nitrogen in MBR

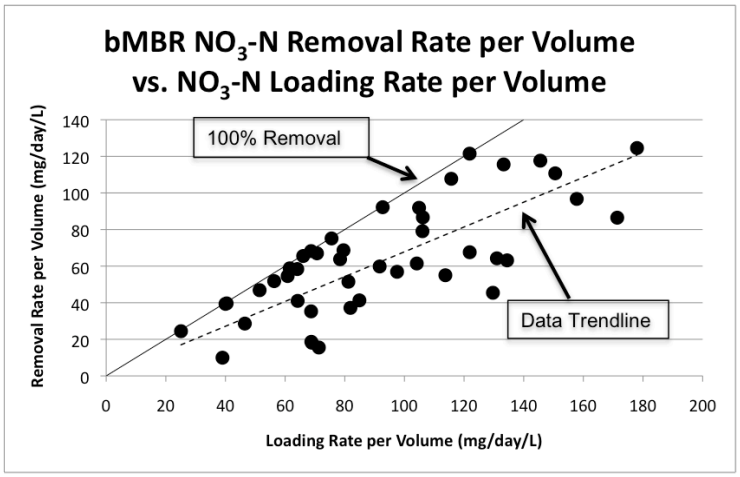


Figure 5.45. Volumetric Removal Rate vs. Loading Rate of Nitrate-Nitrogen in the Anoxic Zone of the bMBR

The nitrate-nitrogen removal rate with respect to loading rate decreased after addition of the larger blower as shown in Figure 5.46. However, this is most likely due to increased loading of nitrate-nitrogen due to more complete

nitrification (Figure 5.39) than a decrease in denitrification rate. The data trend line suggests decreased denitrification as loading increases. The limiting factor in this process is most likely dissolved organic carbon concentration in the influent.

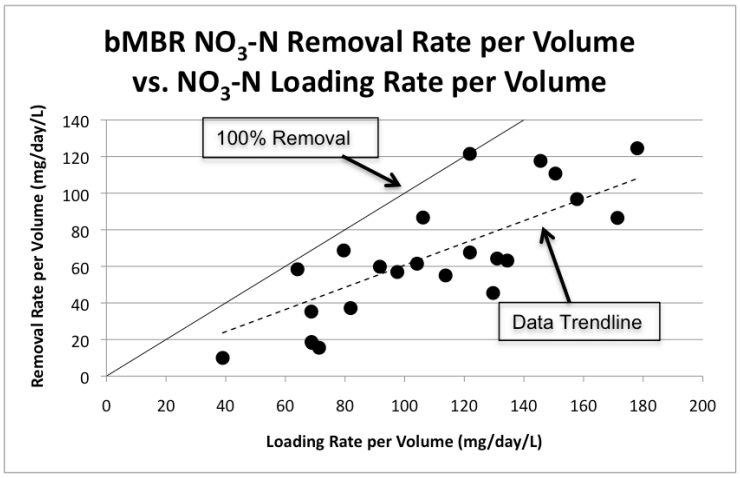


Figure 5.46. Volumetric Removal Rate vs. Loading Rate of Nitrate-Nitrogen in the Anoxic Zone of the bMBR After Large Blower Installation

As previously noted, there was significantly more organic nitrogen in 2009 than in 2010. Figure 5.47 shows Influent total nitrogen concentration for the MBR over the summer of 2009 divided into ammonia-nitrogen and organic nitrogen (ON). Dotted arrows help to illustrate primary pathways of degradation for ammonia and organic nitrogen. Numbers in parentheses are mean influent and effluent concentrations with units of mg/L. Figure 5.48 and 5.49 show the same type of illustration for the bMBR and bMBR after large blower installation, respectively.

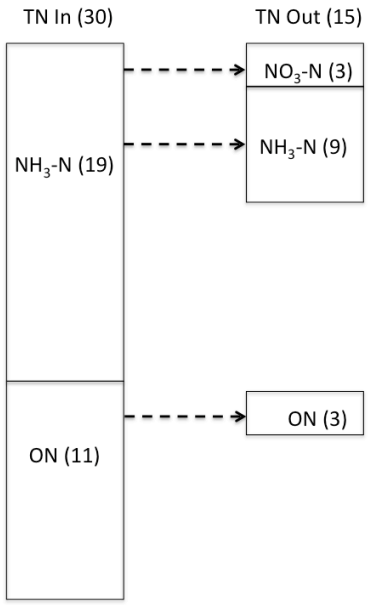


Figure 5.47. Degradation Pathways of Nitrogen Species in the MBR

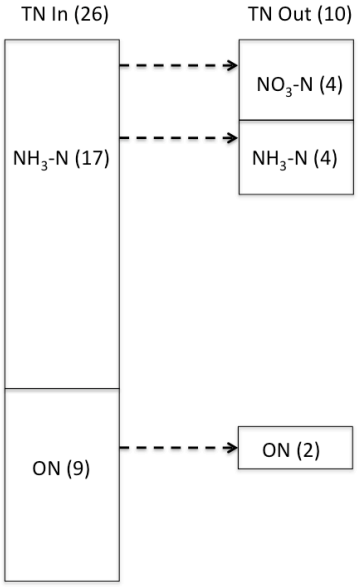


Figure 5.48. Degradation Pathways of Nitrogen Species in the bMBR

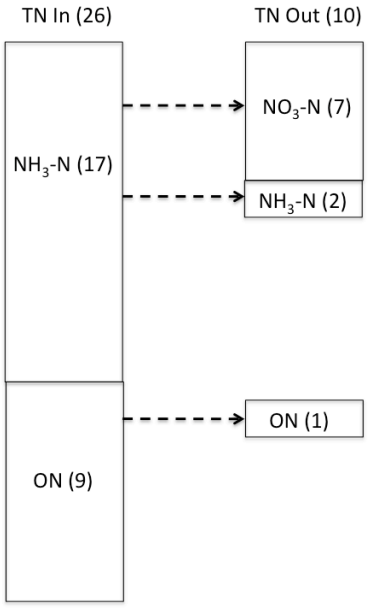


Figure 5.49. Degradation Pathways of Nitrogen Species in the bMBR after Large Blower installation

Comparing Figure 5.43 with Figures 5.44 and 5.45 we can see that the MBR had a greater organic nitrogen concentration than the bMBR. The different ratios of ammonia to organic nitrogen between the MBR and bMBR contribute to the apparent similarity of total nitrogen removal in the comparative analysis.

Table 5.5 shows average volumetric removal for the MBR, the bMBR and the bMBR after the installation of the large blower on July 15. Data for the bMBR was calculated using aerobic volume for ammonia removal and anoxic volume for nitrate removal. Total nitrogen removal between the two reactors is somewhat similar, and one might conclude that there is no significant difference in removal since the variances fall within each other. However, the bMBR had

significantly greater ammonia-nitrogen removal, which is a key step in biological degradation of nitrogen in the wastewater stream.

Table 5.5. Volumetric Removal Comparison per Liter in Reactors

	MBR Volumetric Removal Average	bMBR Volumetric Removal Average	bMBR Volumetric Removal Average (After Large Blower Addition)
COD (g/day/L)	2.0 ± 0.3	1.3 ± 0.2	1.3 ± 0.2
TN (mg/day/L)	81 ± 20	72 ± 9	70 ± 15
NH_3-N (mg/day/L)	51 ± 18	190 ± 30	230 ± 40
NO_3-N (mg/day/L)	38 ± 14	67 ± 12	64 ± 15

The values in Table 5.5 for COD and total nitrogen should be slightly higher for the MBR than the bMBR due to higher influent concentrations in 2009 for the MBR and the greater overall volume of the bMBR. High variance may be partially due to peaks and valleys of influent concentration throughout the testing period (i.e., the reactor can only remove as much of a pollutant as is introduced to the reactor). The larger difference between the reactors with regard to ammonia-nitrogen and nitrate-nitrogen is due to significant differences in active reactor volumes used for the calculation of these values. While the MBR had a higher concentration of BOD/COD, which would presumably have increased

denitrification rates in the bMBR, the mean temperature was one degree Celsius higher in 2010, which would have increased biological degradation rates for the bMBR.

Table 5.6 shows percentage of volumetric removal per volumetric loading for the MBR, bMBR and the bMBR after larger blower installation.

Table 5.6. Percentage of Constituent Removed Comparison

	MBR Percentage Removal	bMBR Percentage Removal	bMBR Percentage Removal (After Large Blower Addition)
COD (%)	95 ± 1	94 ± 1	96 ± 1
TN (%)	53 ± 7	58 ± 4	56 ± 6
NH_3-N (%)	58 ± 16	79 ± 6	93 ± 4
NO_3-N (%)	74 ± 10	71 ± 7	57 ± 9

Table 5.6 helps to give perspective on reactor performance with respect to influent load or, in the case of nitrate, effectiveness of nitrification as well as differences between volumes of the reactors. COD serves as an excellent example of this. Even though volumetric removal of COD was significantly higher in the MBR due to higher influent concentrations and smaller reactor volume, the percentage removal difference between the reactors is not significant

– presumably due to the physical separation of solids from liquid. Analysis of total nitrogen, however, is much more difficult. Off hand it appears that there is no significant difference in total nitrogen removal between the reactors. However, referring back to Tables 5.2 and 5.3, average influent and effluent values for total nitrogen and nitrogen species suggests that the bMBR was significantly more effective at removing total nitrogen and ammonia-nitrogen. It may be that this method of analyzing total nitrogen removal isn't valid because of the variety of nitrogen species that can contribute to the total nitrogen concentration. For example, while this analysis takes loading into account, it doesn't account for the ratio of organic nitrogen to ammonia-nitrogen. Tables 5.2 and 5.3 suggest that the concentration of organic nitrogen was roughly 20% higher in 2009. If both reactors are capable of removing organic particles via membrane filtration than an objective analysis of total nitrogen removal may not be possible unless the reactors are tested under laboratory conditions.

While maintaining D.O. at or above 2 mg/L in the aerobic zone increased ammonia-nitrogen removal from 79.1% to 93.1%, nitrate-nitrogen appears to have decreased. While the decrease in total nitrogen removal after the large blower was installed falls within the margin of error, the percentage of nitrate removed is decidedly less. However, this is due to increased loading of nitrate on the system rather than a decrease in reactor efficacy. At such high loading rates an external carbon source may be need to fully denitrify the liquor, or the internal return ratio should be lowered to increase the food to microorganism (F/M) ratio.

The bMBR used approximately 230 watts in additional power over the MBR. This translates into a 1-watt increase for every 0.13 mg/day/L increase in total nitrogen removal, assuming high variance for the bMBR and low variance for the MBR. Based strictly on the mean among the data, this translates into 0.02 mg/L*day of total nitrogen removed for every additional watt used. However, if the mixer was more precisely matched with the power needed for mixing and hydraulic lift, than less power may be needed to effectively operate the system.

Table 5.6 displays a system comparison that was first visited in the literature review. The last two rows display information on the bMBR as operated throughout the summer, and after the installation of the large blower on July 15.

Table 5.7. Process Performance Comparisons Between bMBR and Other Systems (adapted from Patel 2005)

Process	**COD:TKN Ratio**	**HRT (h)**	**SRT (day)**	**Eff. COD (mg/L)**	**Eff. NH_4-N (mg/L)**	**Eff. NO_3-N (mg/L)**	**Eff. TKN (mg/L)**
A2O	100:13.6	10	10	–	7.5	4.5	10.5
UCT	100:10.7	17.5	20	29	–	9.6	4.0
MUCT	100:9.4	17.5	20	35	–	9.3	4.8
MBR (Single Stage)	100:12	7.5	20	17	<0.3	43	<0.5
Modified MBR	100:9.4	12	20	3	0.4	5.8	–
bMBR	100:8.5	5.9	41	14	4.0	4.3	5.7
bMBR (after fine-bubble addition)	100:8.9	5.9	41	11	1.3	6.8	3.2

Based on the comparative information in Table 5.6, it appears that the bMBR is competitive with other technologies in terms of its effectiveness. Efforts were made to compare these, and other, reactors to the bMBR and MBR based on volumetric removal of COD, TN, NH_3-N and NO_3-N, but the data necessary to calculate the volumetric removal and loading rates was not readily available.

6. CONCLUSIONS AND RECOMMENDATIONS

6.1. CONCLUSIONS

The goal of this research was to demonstrate the efficacy of a unique MBR design meant to improve nitrogen removal over a conventional design while reducing energy inputs generally considered necessary for conventional systems with a pre-anoxic zone. Results and conclusions in Chapter 5 suggest that this goal was met, even if some objectives were not fully achieved.

The first “conventional” MBR was designed, fabricated, and operated, completing the first objective – design, build and operate a conventional-type MBR – outlined in Chapter 3. The configuration of the MBR test throughout 2009 may have been detrimental to its performance. Having the coarse-bubble diffusers elevated above the bottom of the reactor allowed for a significant amount of biomass settling. Additional diffusers, a smaller volume, or greater mixing in general may have increased reaction rates and resulted in more efficient constituent removal. That being said, it is intuitive that a reactor configuration with defined zones between anoxic and aerobic conditions should provide more consistent treatment than reactors where these zones are intermingled.

Objectives two and four, which were to design and construct a bMBR and to compare the reactors, respectively, were fully completed. Improvements to

controls and operation of the “conventional” MBR resulted in more reliable operation of the bMBR, with the exception of increased fouling observed during bMBR operation. The bMBR was successfully operated without the use of a pump to return liquor to the pre-anoxic zone. Once steady-state was reached and ideal operational parameters established, the reactor regularly demonstrated the ability to produce effluent that meets US EPA requirements for losing streams without requiring organic carbon addition. During these periods there were occasional releases of nitrate-nitrogen in concentrations higher than 10 mg/L. However, these seemed to be the exception rather than the norm, and were likely the result of high loading of nitrate-nitrogen onto the system by way of ammonia-nitrogen oxidation. Comparison of the reactors based on volumetric removal of nitrogen species, as well as removal rate as a function of loading rate, showed that the bMBR was more effective at removing ammonia-nitrogen and may be more effective at removing total nitrogen and nitrate-nitrogen.

Objective three, operating the reactors over the same timeframe, was partially met. Although both reactors were operated over a similar period of time and during similar seasons, operating the reactors under the same conditions was all but impossible. Ideally, the two reactors would have operated side-by-side on the same influent. However, limitations on membrane acquisition nullified that possibility. Differences in temperature and influent characteristics could have played a role in differences among operational results of the reactors. It is widely accepted that the wastewater generated from the surrounding community used in these experiments is generally dilute compared to typical

values of municipal wastewater. The BOD/COD:TN ratio seemed to be lower in 2010 than it was in 2009, with a higher ratio being more conducive to denitrification. Had the MBR/bMBR experiment been run over the same period of time and under the same conditions a greater understanding of their differences could be obtained.

Objective five was only partially completed. Although no measurements were taken of energy use during operation, it is assumed that the bMBR required the same amount of energy as the MBR with the exception of the anoxic mixer and aeration modification. The need for a larger blower and additional fine-bubble diffusion in the bMBR was likely the result of higher and better distributed concentrations of biomass within the reactor, which in turn were the result of better mixing and higher operational temperature. Furthermore, a larger blower for coarse-bubble diffusion may have not been necessary if the influent contained less fat, oil and grease as was indicated by Allen McNece, Plant Manager. According to McNece, fat, oil and grease concentrations in the influent were much greater in 2010 than in 2009. Therefore, differences in influent characteristics again precluded us from making an accurate comparison.

Had the influent been consistent between 2009 and 2010, than the only difference in energy consumption may have been the mixer. The mixer used 230 watts in post-experiment testing, and a translation of mg/L*day per additional watt of power was made. However, since energy consumption was not measured throughout the testing period we can’t make a firm determination. Furthermore, the mixing motor was not carefully selected for the bMBR and it

could be that a smaller or more efficient motor would have sufficed. Ultimately, the removal per additional power required comparison is only applicable between reactors of similar size. Scaling would almost certainly make a difference in this comparison.

Objective six, providing an overall assessment, is only possible with the help of significant assumptions – that differences in influent, environment, and, in the case of the MBR, mixing would have had no affect on reactor performance. With these assumptions we can declare the bMBR to be a superior design over the MBR based on its ability to remove typical wastewater constituents that are typically regulated by operating permits. While it is probable that the MBR would have performed better with increased mixing and higher temperatures, it is also possible that the bMBR can perform better than what this experiment suggests. Adjusting airflow to the fine-bubble diffusers could better control proper distribution of D.O. concentration in the aerobic zone and anoxic zone. Should the bMBR reactor be sized-up and put into operation full-time, the system would need telemetry technology and PLC control along with an aerobic D.O. probe to constantly monitor D.O. concentration and adjust flow rate as needed to maximize reactor effectiveness.

The overall assessment of these reactors based strictly on volumetric removal or percentage of volumetric removal per volumetric loading is misleading in real-world applications. Few municipalities across the U.S. limit discharges of total nitrogen. Those that do may set the TKN (not total) limit as low as 10-15 mg/L, with TKN being total nitrogen minus nitrate/nitrite-nitrogen concentrations.

Ammonia-nitrogen concentrations are typically set at 2-3 mg/L, with nitrate/nitrite-nitrogen concentrations set at 10-15 mg/L when issued. Considering these restrictions on wastewater discharge, the bMBR after larger blower installation is the only one of the three configurations that may be viable. The MBR was not able to provide ample nitrification, which is the first step in nitrogen removal due to the eutrophication potential of ammonia once it enters the environment. Even the bMBR after large blower installation doesn't meet the discharge requirements if the variances are considered, which they are for pilot-scale testing of new reactors at a confidence interval of 98% rather than the 95% that was used in this study. Therefore, the ultimate conclusion of this research is that the bMBR after large blower installation is the only viable reactor tested in this study. With better environmental and process controls it is possible that even the most stringent nitrogen regulations could be met.

6.2 RECOMMENDATIONS

This experiment should be repeated with three reactors run off the same influent stream in parallel. The first reactor would be a MBR with the same volume as the aerobic zone in the bMBR. The second would be a MBR with the same volume as that of the bMBR. A bMBR would constitute the third reactor. Such an arrangement would provide better apples-to-apples comparison of different reactor configurations under the same conditions. At the same time, the

limits of denitrification may be tested using methanol for complete nitrate-nitrogen removal. It may be necessary to operate these reactors in the lab under controlled conditions to ensure standardization.

If this first experiment demonstrated the bMBR to be a superior design, a second experiment should be carried out with a bMBR as designed in 2010 and a second bMBR using an appropriately sized pump to return liquor to the pre-anoxic zone. Energy consumption comparisons should be made to provide estimated cost-savings between the two configurations. Also, additional configurations may be used by applying the same principle and creating an anaerobic zone before the anoxic and aerobic zones to encourage PAO growth and biological phosphorous removal.

Once the second experiment is complete, a pilot-scale experiment should be carried out to help understand the implications and challenges of scaling the bMBR. Assessing the applicability of this reactor to future installations and requirements is key in forming the foundation for widespread acceptance.

APPENDIX

Raw Data and Analysis

Included with this thesis is a CD-ROM, which contains the Excel documents with data and calculations for the MBR and bMBR. Both files are formatted in Excel 98.

BIBLIOGRAPHY

Ahmed, Z. C., J. Lim, B.R. Song, K.G. Ahn, K.H. (2007). "Effects of Sludge Retention Time of Membrane Fouling and Microbiological Community Structure in a Membrane Bioreactor." Journal of Membrane Science **287**(2): 211-218.

Berenson, M., Levine, David M. (1999). Basic Business Statistics: Concepts and Applications. New Jersey, Prentice-Hall, Inc.

Brannock, M., G. Leslie, et al. (2009). "Optimising mixing and nutrient removal in membrane bioreactors: CFD modelling and experimental validation." Desalination: 815-818.

Buer, T. and J. Cumin (2008). "MBR module design and operation." Desalination **250**(3): 1073-1077.

Chae, S. R., Kang, S.T., Watanabe, Y., Shin, H.S. (2006). "Development of an Innovative Vertical Submerged Membrane Bioreactor (VSMBR) for Simultaneous Removal of Organic Matter and Nutrients." Water Research **40**: 2161-2167.

Chae, S. R. A., Y.T. Kang, S.T. Shin, H.S. (2006). "Mitigated Membrane Fouling in a Vertical Submerged Membrane Bioreactor (VSMBR)." Journal of Membrane Science **280**(1-2): 572-581.

Cho, J. S., K.G. Hyup Lee, S. Ahn, K.H. (2005). "Sequencing Anoxic/Anaerobic Membrane Bioreactor (SAM) Pilot Plant for Advanced Wastewater Treatment." Desalination **178**(1-3): 219-225.

Daigger, G. T. R., Bruce E. Adham, Samer Andreottola, Gianni (2005). "Are Membrane Bioreactors Ready for Widespread Application?" Environmental Science and Technology **39**(19): 399A-406A.

EPA (1999). "EPA Method 160.2." Retrieved 4/11, 2011, from http://www.epa.gov/region9/qa/pdfs/160_2.pdf.

Field, R. W. W., D. Howell, J.A. Gupta, B.B. (1995). "Critical Flux Concept for Microfiltration Fouling." Journal of Membrane Science **100**: 241-248.

Fleischer, E. J., Broderick, Thomas A., Daigger, Glen T., Fonseca, Anabela D., Holbrook, David R., Murthy, Sudhir N. (2005). "Evaluation of Membrane Bioreactor Process Capabilities to Meet Stringent Effluent Nutrient Discharge Requirements." Water Environment Research **77**(2): 162-178.

Gayle, B. P. (1989). "Biological Denitrification of Water." Journal of Environmental Engineering **115**.

Grady Jr., C. P. L., G. T. Daigger, et al. (1999). Biological Wastewater Treatment. New York, Marcel Dekker, Inc.

Han, S. S. B., T.H. Jang, G.G. Tak, T.M. (2005). "Influence of Sludge Retention Time on Membrane Fouling and Bioactivities in Membrane Bioreactor System." Process Biochemistry **40**: 2393-2400.

Hasar, H. (2009). "Simultaneous removal of organic matter and nitrogen compounds by combining a membrane bioreactor and a membrane biofilm reactor." Bioresource Technology **100**(10): 2699-2705.

Howarth, R. W. and R. Marino (2006). "Nitrogen as the limiting nutrient for eutrophication in coastal marine ecosystems: Evolving views over three decades." Limnology and Oceanography **51**(1 II): 364-376.

Jiang, T. K., Maria D. van der Meer, Walter G.J. Vanrolleghem, Peter A. Schippers, Jan C. (2003). "The Role of Blocking and Cake Filtration in MBR Fouling." Desalination **157**(1-3): 335-343.

Judd, S. J., Claire (2006). The MBR Book: Principles and Applications of Membrane Bioreactors in Water and Wastewater Treatment. Kidlington, Oxford, UK, Elsevier, Ltd.

Kimura, K., M. Enomoto, et al. (2007). Modification of submerged membrane bioreactors (MBRs) by inserting baffles: Pilot scale study. Water Science and Technology. **55:** 119-126.

Kimura, K., R. Nishisako, et al. (2008). "Baffled membrane bioreactor (BMBR) for efficient nutrient removal from municipal wastewater." Water Research **42**(3): 625-632.

Kimura, K. and Y. Watanabe (2005). Baffled membrane bioreactor (BMBR) for advanced wastewater treatment: Easy modification of existing MBRs for efficient nutrient removal. Water Science and Technology. **52:** 427-434.

KOCH Membrane Systems (2008). Main Dimension Puron – Module 1.3m^2. M. Fitch.

Le Clech, P., Jefferson, Bruce, Chang, In Soung, Judd, Simon J. (2003). "Critical Flux Determination by the Flux-Step Method in a Submerged Membrane Bioreactor." Journal of Membrane Science **227**: 81-93.

Le-Cletch, P. C., Vicki Fane, Tony A.G. (2006). "Fouling in Membrane Bioreactors Used in Wastewater Treatment." Journal of Membrane Science **284**(1-2): 17-53.

Liang, S. L., C. Song, L. (2007). "Soluble Microbial Products in Membrane Bioreactor Operation: Behaviors, Characteristics and Fouling Potential." Water Research **41**(1): 95-101.

LMNO Engineering, R., and Software, Ltd. (2007). "V-Notch (Triangular) Weir Calculator: Discharge and Head Calculations, Equations, and Guidelines for Water Flow Measurement in Streams and Channels." Retrieved November, 2009, from www.lmnoeng.com/Weirs/vweir.htm.

Loftus, T. (2003). "BOD Test Requirements." Lagoon Systems In Maine. Retrieved 4/11, 2011, from http://www.lagoonsonline.com/laboratory-articles/bodtest.htm.

Loftus, T. (2003). "Total Suspended Solids." Lagoon Systems in Maine. Retrieved 4/12, 2011, from http://www.lagoonsonline.com/laboratory-articles/tss.htm.

Melin, T. J., B. Bixio, D. Thoeye, C. De Wilde, W. De Koning, J. van der Graaf, J. Wintgens, T. (2006). "Membrane Bioreactor Technology for Wastewater Treatment and Reuse." Desalination **187**(1-3): 271-282.

Meng, F., S. R. Chae, et al. (2009). "Recent advances in membrane bioreactors (MBRs): Membrane fouling and membrane material." Water Research **43**(6): 1489-1512.

Meng, F. S., B. Yang, F. Zhang, H. (2007). "Effect of Hydraulic Retention Time on Membrane Fouling and Biomass Characteristics in Submerged Membrane Bioreactors." Bioprocess and Biosystems Engineering **30**: 359-367.

Metcalf & Eddy, I., G. Tchobanoglous, et al. (2003). Wastewater Engineering, Treatment and Reuse. New York, McGraw-Hill.

Ng, H. Y. T., T.W. Ong, S.L. Toh, C.A. Loo, Z.P. (2006). "Effects of Solid Retention Time of the Performance of Submerged Anoxic/Oxic Membrane Bioreactor." Water Sciences and Technology **53**(6): 7-13.

Ngo, H. H. and W. Guo (2009). "Membrane fouling control and enhanced phosphorus removal in an aerated submerged membrane bioreactor using modified green bioflocculant." Bioresource Technology **100**(18): 4289-4291.

Nywening, J. P. and H. Zhou (2009). "Influence of filtration conditions on membrane fouling and scouring aeration effectiveness in submerged membrane bioreactors to treat municipal wastewater." Water Research **43**(14): 3548-3558.

Park, H. D., Y. H. Lee, et al. (2010). "Reduction of membrane fouling by simultaneous upward and downward air sparging in a pilot-scale submerged membrane bioreactor treating municipal wastewater." Desalination **251**(1-3): 75-82.

Patel, J. N., George Margaritis, Argyrios (2005). "Optimization of Biological Nutrient Removal in a Membrane Bioreactor System." Journal of Environmental Engineering **ASCE/July**: 1021-1029.

Payne, W. J. (1981). Denitrification. New York, Wiley.

Phattaranawik, J. F., A.G. Pasquier, A.C.S. Bing, W. (2007). "Membrane Bioreactor with Bubble-Size Transformer: Design and Fouling Control." AICHE Journal **53**: 243-248.

Psoch, C. S., S. (2006a). "Anti-Fouling Application of Air Sparaging and Backflushing for MBR." Journal of Membrane Science **283**(1-2): 273-280.

Sarioglu, M., G. Insel, et al. (2009). "Model evaluation of simultaneous nitrification and denitrification in a membrane bioreactor operated without an anoxic reactor." Journal of Membrane Science **337**(1-2): 17-27.

Song, K.-G., Kim, Yuri, Ahn, Kyu-Hong (2008). "Effect of Coagulant Addition of Membrane Fouling and Nutrient Removal in a Submerged Membrane Bioreactor." Desalination **221**: 467-474.

Stephenson, T. J., Simon Jefferson, Bruce Brindle, Keith (2000). Membrane Bioreactors for Wastewater Treatment. Padstow, Cornwall, UK, TJ International.

Tchobanoglous, G., F. L. Burton, et al., Eds. (2003). Wastewater Engineering, Treatment and Reuse. New York, McGraw-Hill.

Trussell, R. S. M., R.P. Hermanowicz, S.W. Jenkins, D. (2007). "Influence of Mixed Liqour Properties and Aeration Intensity on Membrane Fouling in a Submerged Membrane Bioreactor at High Mixed Liqour Suspended Solids Concentrations." Water Research **41**(5): 947-958.

Wang, Z. W., Zhichao Yin, Xing Tian, Lumei (2008). "Membrane Fouling in a Submerged Membrane Bioreactor (MBR) Under Sub-Critical Flux Operation: Membrane Foulant and Gel Layer Characterization." Journal of Membrane Science **325**: 238-244.

Werzernak, C. T., and Gannon, J.J. (1967). "Oxygen-Nitrogen Relationships in Autotrophic Nitrification." Applied Microbiology **15**: 1211-1215.

Wisconsin Department of Natural Resources (2003). "Nitrate." Retrieved July 15, 2010, from http://dnr.wi.gov/org/water/dwg/nitrate.htm.

Yang, S., Yang, Fenglin, Fu, Zhimin, Wang, Tao, Lei, Ruibo (2009). "Simultaneous Nitrogen and Phosphorous Removal by a Novel Sequencing Batch Moving Bed Membrane Bioreactor for Wastewater Treatment." Journal of Hazardous Materials **175**: 551-557.

Yigit, N. O. C., G. Harman, I. Koseoglu, H. Kitis, M. (2008). "Effects of Various Backwash Scenarios on Membrane Fouling in a Membrane Bioreactor." Desalination **237**: 346-356.

Zhang, J. S. C., C.H. Zhou, J.T. Fane, A.G. (2006). "Effect of Sludge Retention Time on Membrane Bio-Fouling Intensity in a Submerged Membrane Bioreactor." Seperation Science and Technology **41**(7): 1313-1329.

Zhang, Y., J. Zhou, et al. (2009). "An innovative membrane bioreactor and packed-bed biofilm reactor combined system for shortcut nitrification-denitrification." Journal of Environmental Sciences **21**(5): 568-574.

VITA

Timothy Harrison Canter was born in St. Louis, MO on July 17, 1978. In December 2003 he received his B.S. in Advertising and Marketing from Missouri State University, Springfield. He entered the foodservice industry by accepting positions as The Winnie Group in Springfield, MO followed by Esrock Partners – outside of Chicago – through 2008. Employment with the former allowed him the opportunity to help build Wild Bear Log Homes, a Katahdin Cedar Log Home dealership and subsidiary of The Winnie Group. He continued his career in foodservice marketing by moving to Chicago in September of 2007 to work for Esrock Partners in Orland Park, IL.

By 2008 he had moved back to Springfield, MO and continued to work through prerequisites for the Masters program in Environmental Engineering at Missouri University of Science and Technology. He enrolled for the first semester of the Masters program in January of 2009. By September of 2009 he had begun to assist Frontier Environmental Technology with operational functions including marketing, sales and expansion strategy.

Tim is involved in the Missouri Stream Team program offered through Missouri Department of Conservation, and is also a member of the Missouri Conservation Federation. Tim is a member of Water Environment Federation and Missouri Water Environment Association, where he currently sits on the Small Flows Committee.